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TECHNO-ECONOMIC EVALUATION OF THIOSALT TREATMENT PROCESSES

by

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SUMMARY

Research aimed at the control of thiosalts generated during base metal milling has been conducted by Environment Canada, CANMET, Brunswick Mining and Noranda Inc., for the past decade. Measures studied to reduce the generation of thiosalts were found to generally lower metal recoveries. Subsequently, most of the work consisted in developing methods for the destruction of thiosalts in mill effluents. Since available cost estimates lacked a uniform basis, the current project was undertaken with the aim of deriving flowsheets and consistent cost estimates for the treatment of thiosalt effluents from a 10,000 t/d mine/mill complex. Three situations were considered:

- The base case, with an effluent flow of 720 m^3/h and 24,500 kg/d thiosalt discharge.
- Reduced thiosalt loading, with an effluent flow of 720 m^3/h, and a 12,300 kg/d thiosalt discharge (100% flow, 50% thiosalts).
- Reduced flow and thiosalt loading, with an effluent flow of 360 m^3/h, and a 12,300 kg/d thiosalt discharge (50% flow, 50% thiosalts).

The following methods were evaluated:

- Air Oxidation (biological, catalyzed with polyvinyl pyridine-copper, catalyzed with copper sulphide)
- Chemical Oxidation (H_2O_2, Cl_2, O_3, SO_2 - air)

- Membrane and Electrochemical processes
- Other (reduction by iron, alkaline pressure oxidation, activated carbon adsorption, disposal to the sea, natural degradation)

The capital and operating plus maintenance costs derived are presented in the summary table (page 5), that also indicates the overall annual costs (annualized capital, operating and maintenance costs). As shown, the capital costs vary from $8.9 million to $79 million for the base case, and from $7.0 million to $45.3 million for reduced flow/thiosalts discharge. The total costs for the capitalization and operation of a thiosalt treatment facility range from $3.3 million to $70 million per year for the base case, and from $2.2 million to $36.3 million per year for the reduced flow/thiosalts discharge. The lowest cost options include:

- Natural Degradation
- Biological Air Oxidation
- Disposal to the Sea
- Copper/Sulphide Catalyzed Air Oxidation

These methods involve, however, a high degree of technical uncertainty, as follows:

- Natural Degradation

- Lengthy tests would be required to test the concept and to determine actual retention time requirements. In many areas, the extensive pond surface required may not be available, or the earthworks may be very expensive because of local topology.

- Biological Treatment

  - Heavy metals (especially lead) have been shown to interfere with biological activity.

  - The process requires extensive start-up times (one to two months). Interruptions in the process (e.g. strikes, mill down-time) would require lengthy restarting, during which unacceptable effluents would be released unless other methods are used concurrently (e.g. full-scale chemical oxidation).

  - Although the destruction of thiosalts using bacteria does not result in the generation of voluminous biological solids, some solids may form (e.g. sulphur). Insufficient information exists to determine potential detrimental effects ver a pond life of 15 to 25 years.

  - The biological treatment flowsheet is based on pilot tests of 1.2 to 1.8 m^3/h. Full scale facilities would require very significant (and highly uncertain) scaling-up by a factor of 600.

- Disposal to the Sea

  - Flow of lime-treated effluents through a pipeline will result in scaling. It is not known whether routine maintenance, or the use of scale inhibiting reagents will be sufficient to control this problem.

  - A leak in the pipeline would result in the discharge of thiosalts to the environment, and possible acid generation.

- Copper Sulphide Catalyzed Air Oxidation

Only laboratory scale tests have been conducted to date to evaluate this technology. Although the results are promising, the estimated costs are still relatively high. Extensive test work would be required (3 to 5 years) to evaluate the technical feasibility of this method for full-scale treatment.

The methods which could be implemented more readily are those utilizing chemical oxidants (e.g. H_2O_2, Cl_2) where the technology is currently used by the mining industry for the treatment of cyanide effluents. Overall costs for the capitalization and operation of such thiosalt facilities would be high, with costs of \$16 to \$19 million for the base case, and \$8.6 to \$10 million for reduced thiosalts/flow.

Thus, as indicated above, all the costs are very high, and none of the methods evaluated has been proven to be consistently effective for large-scale treatment of actual effluents.

THIOSALT TREATMENT ALTERNATIVES - SUMMARY OF COSTS

		OVERALL CAPITAL (MILLIONS $)			ANNUAL OPERATING[1.] AND MAINTENANCE (MILLIONS $)			OVERALL ANNUAL COSTS (CAPITALIZATION OPERATING MAINTENANCE) (MILLIONS $)		
		Base Case[2.]	100% Flow 50% S_2O_3	50% Flow 50% S_2O_3	Base Case[2.]	100% Flow 50% S_2O_3	50% Flow 50% S_2O_3	Base Case[2.]	100% Flow 50% S_2O_3	50% Flow 50% S_2O_3
1. Biological Air Oxidation	1.1 Without post-treatment	13.5	10.1	9.3	2.5	1.5	1.5	5.2	3.5	3.4
	1.2 With H_2O_2 post-treatment	15.6	11.7	10.9	4.0	2.3	2.2	7.1	4.6	4.4
2. Catalyzed Air Oxidation	2.1 With Polyvinyl Pyridine Copper	32.2	20.2	19.4	19.6	10.6	10.1	26.0	14.6	14.0
	2.2 With Converter White Metal (copper sulphide)	14.8	11.0	9.7	4.2	3.1	2.4	7.2	5.3	4.5
3. H_2O_2 Oxidation		8.9	8.0	7.0	14.9	7.9	7.7	16.7	9.5	9.1
4. Ozone Oxidation		50.5	30.7	29.1	8.4	4.8	4.8	18.5	11.0	10.6
5. Chlorine Oxidation	5.1 Purchased Lime	10.0	8.2	7.1	18.4	9.5	9.4	20.4	11.1	10.8
	5.2 On-Site Lime Production	17.2			16.2			19.6		
6. SO_2 - Air Oxidation	6.1 Sulphur/Purchased Lime	13.3	10.6	9.3	8.6	4.6	4.5	11.3	6.8	6.4
	6.2 Sulphur/On-Site Lime Prod.	22.0			5.8			10.2		
	6.3 Pyrite/Purchased Lime	15.5	12.4	11.1	7.3	4.0	4.0	10.4	6.5	6.2
	6.4 Pyrite/On-Site Lime Prod.	24.2			4.5			9.3		
7. Electrochemical Oxidation		79.0	55.1	45.3	17.1	11.0	8.9	32.9	22.1	17.9
8. Reverse Osmosis	8.1 Brine Treatment by Evaporation	22.2	22.2	14.5	10.6	10.6	5.7	15.1	15.1	8.6
	8.2 Electrochemical Brine Treatment	44.7	35.4	26.9	12.4	9.4	6.6	21.3	16.5	12.0
	8.3 Alkaline Pressure Oxidation Brine Treatment	33.5	30.9	22.5	11.7	10.2	6.3	18.4	16.4	10.8
9. Alkaline Pressure Oxidation		23.7	21.0	16.3	12.5	11.2	6.5	17.3	15.4	9.7
10. Reduction by Iron	10.1 With H_2O_2 Treatment of H_2S	13.4	12.7	9.3	14.7	12.2	7.5	17.4	14.8	9.3
	10.2 With Claus Reactor	16.9			13.7			17.2		
11. Activated Carbon Adsorption	11.1 Thermal Carbon Regeneration	37.0	25.9	24.1	65.2	33.1	33.0	72.6	38.3	38.9
	11.2 Alkaline Carbon Regeneration/Electrochemical Oxidation of Alkali	23.6	17.0	15.2	9.1	5.8	4.9	13.8	9.2	8.0
	11.3 Alkaline Carbon Regeneration/Alkaline Pressure Oxidation of Alkali	19.1	14.5	12.7	7.8	5.0	4.3	11.7	7.9	6.8
12. Disposal to the Sea (Pipeline)		28.9	28.7	24.5	1.4	1.3	1.1	7.2	7.1	6.0
13. Natural Degradation		12.4	11.8	9.4	1.6	1.1	1.0	4.1	3.4	2.9

1. Exclusive of capital charges.
2. Base Case = 720 m^3/h flow, 24,500 kg/d thiosalts.

1. INTRODUCTION

The milling of base metal sulphide ores often results in the generation of partially oxidized sulphur compounds (e.g. $S_2O_3^{2-}$, $S_3O_6^{2-}$ and $S_4O_6^{2-}$), known as thiosalts. These anions are discharged in mill effluents, and are generally unaffected by conventional lime treatment. Subsequently, on entering receiving waters, thiosalts are oxidized to sulphates, resulting in pH depression. Actual acidity levels are determined by such factors as dilution, temperature, thiosalt concentrations, and alkaline content.

The occurrence and treatment of thiosalts in milling effluents has been researched intensively by CANMET, Environment Canada, and Noranda Inc., among others. Sampling and analytical techniques for the determination of thiosalts in milling effluents were developed. Sources of thiosalts were investigated, and generation was found to be related mainly to:

- pH of the milling circuit;
- sulphide content of the ore;
- residence time during flotation;
- temperature.

Methods for the reduction of thiosalt generation during the milling process were evaluated, but were shown to result in lower metal recoveries. Major efforts have therefore been expended in defining treatment methods for the destruction of thiosalts in mill effluents. Most of these methods were evaluated in preliminary feasibility studies based on information or laboratory scale tests. Biological treatment of thiosalts had indicated most promise, and on-site pilot tests were conducted at Brunswick Mining and Smelting, Mining Division, in Bathurst, N.B.

Although estimates of operating and capital costs were determined for some of the thiosalt treatment alternatives, there was a lack of consistency regarding such parameters as thiosalt loadings, effluent flow, pre- and post-treatment facilities, and unit costs. Consequently, CANMET's Mineral Sciences Laboratories sponsored this techno-economic study aimed at defining flowsheets for full-scale thiosalt treatment facilities, and providing realistic operating and capital cost estimates on a consistent basis. The project was undertaken jointly by the Noranda Research Centre, and Gionet, Mellor, Liebich Associates, and the results are presented in this report.

2. PROJECT METHODOLOGY AND BASIS

2.1 Mine and Effluent Parameters

Thiosalt treatment facilities evaluated in this study are based on the mill parameters of Brunswick Mining and Smelting Corporation, (BMS). The company operates an underground mine and concentrator complex near Bathurst in north-eastern New Brunswick, with an extraction and milling capacity of 10,000 t/d ore containing zinc, lead, copper and silver. Milling consists of wet grinding, pulp aeration with SO_2 addition, flotation, and upgrading with heating. The mill process parameters are summarized in Table I. The sources of thiosalt generation during the milling process are itemized in Table II indicating the main inputs being due to grinding, aeration and flotation operations. The distribution of thiosalt species at BMS is indicated in Table III, and shows that $S_2O_3^{2-}$ and $S_3O_6^{2-}$ accounts for over 90% of the total thiosalts produced.

TABLE I

BMS MILL PROCESS PARAMETERS

Ore milled, t/d	10,000		
Sulphides in ore, %	80		
Grind	68% below 400 mesh		
Fresh water used, m^3/h	640		
Retention of ore in mill, h	3		
Metallurgy	**Cu**	**Pb**	**Zn**
Mill feed grade	0.3	3.0	7.0
Copper concentrate grade, %	20.0	8.0	5.0
Lead concentrate grade, %	0.3	36.0	6.0
Zinc concentrate grade, %	0.2	1.0	52.0
Bulk concentrate grade, %	0.8	21.0	33.0
Recovery, %	40.0	60.0	80.0

TABLE II

GENERATION OF THIOSALTS IN THE BMS MILL CIRCUIT*

Operation	Thiosalts Generated kg/min	% of Total
Adding internal reclaim water to lines 1 & 2	1.4	5.6
Grinding (lines 1, 2 & 3)	3.2	13.2
Aeration (lines 1 & 2)	3.3	13.6
Flotation (lines 1, 2 & 3)	4.3	17.8
Flotation (primary upgrading)	5.2	21.4
Flotation (secondary upgrading)	4.1	17.0
Adding SO_2 (primary upgrading)	0.5	1.9
Adding SO_2 (secondary upgrading)	1.0	4.3
Heating 85°C (primary upgrading)	0.8	3.4
Heating 85°C (secondary upgrading)	0.4	1.8
TOTAL	24.2	100.0

*As determined during September 19-30, 1977(1).

TABLE III

DISTRIBUTION OF THIOSALT SPECIES AT BMS(1)

Thiosalt Species	kg/min	% of Total
Thiosulphate ($S_2O_3^{2-}$)	12	50
Trithionate ($S_3O_6^{2-}$)	10	41
Tetrathionate ($S_4O_6^{2-}$)	2	9
Total Thiosalts	24	100

The BMS water balance is illustrated in Figure 1. As indicated, the main water discharges from the mill are in tailings (1000 m^3/h) and thickener overflows (250 m^3/h). Part of the tailings decant (480 m^3/h) is recycled to the mill, and the remainder discharged. The thiosalt balance (Figure 2) shows a net discharge of 14,500 kg/d S_2O_3 in tailings water, and 10,000 kg/d S_2O_3 in thickener overflows. Approximately 1500 kg/d thiosalts are retained in the tailings as interstitial water.

The BMS tailings decant and thickener overflow water quality parameters are presented in Tables IV and V.

TABLE IV

BMS TAILINGS WATER QUALITY PARAMETERS*
(as mg/L, except for pH)

Parameter	
pH	9.2
Pb	7.8
Zn	6.4
Cu	0.5
Fe	3.7
Ca	395
Mg	6.5
Na	650
Thiosalts, as S_2O_3	1320
SO_4	1640
Total Water Hardness as $CaCO_3$	1010

*Average 1983 values.

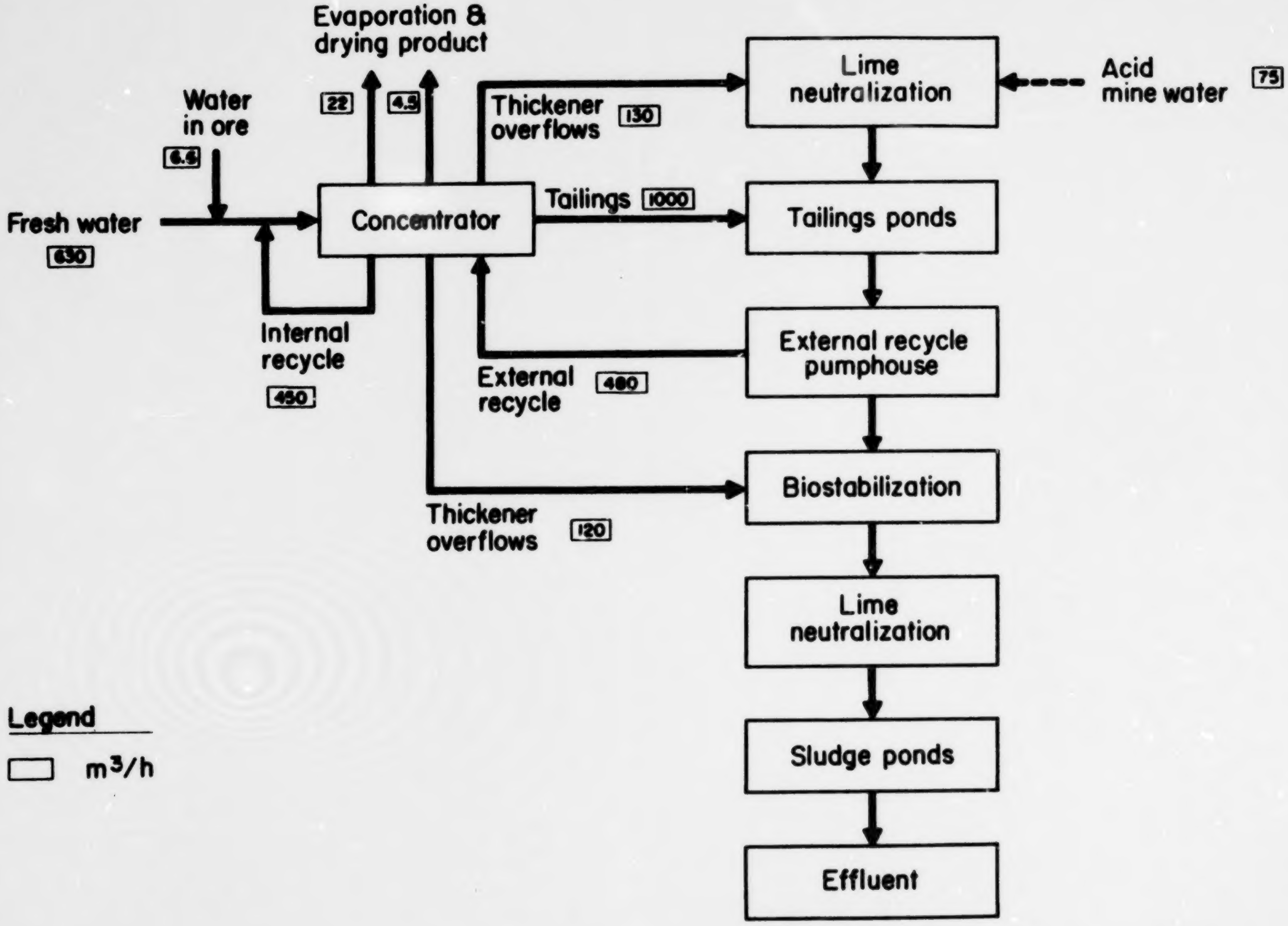

Figure 1 - BMS Water Balance

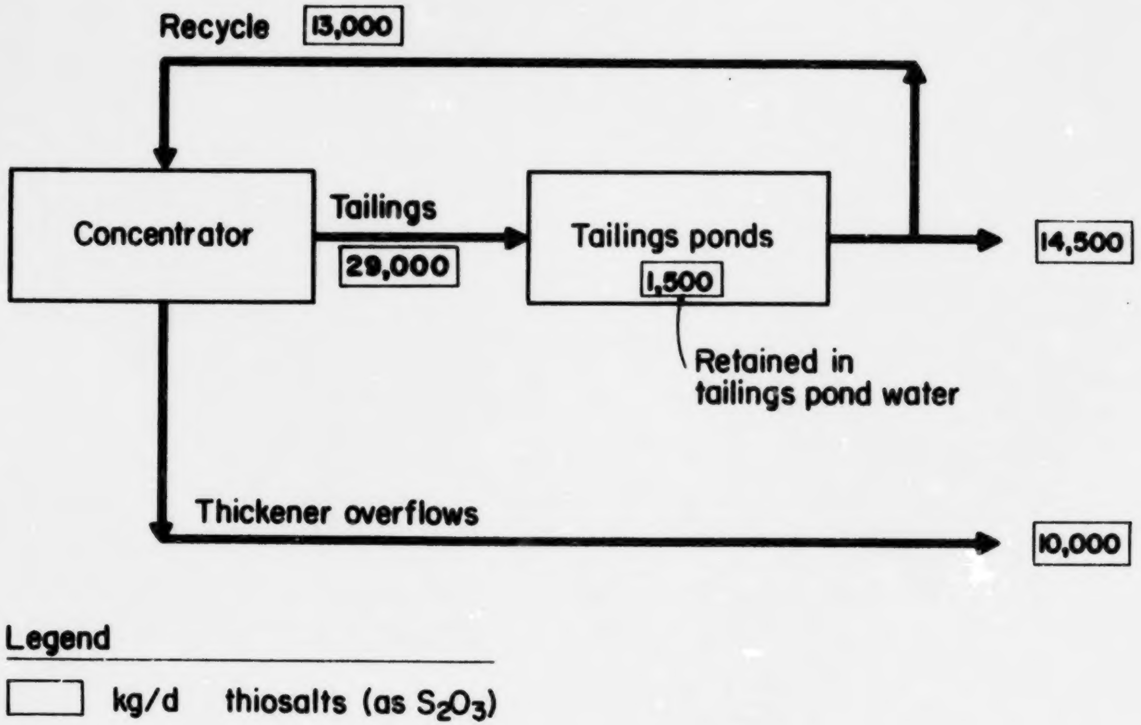

Figure 2 - BMS Thiosalt Mass Balance

TABLE V

BMS THICKENER OVERFLOWS - WATER QUALITY PARAMETERS[1]

(as mg/L, except pH, temperature, and flow)

Parameter	Lead Thickener Overflow	Zinc Thickener Overflow	Copper Thickener Overflow	Bulk Thickener Overflow	Bulk Cooling Overflow	Total (weighted average)
Flow[2] (L/min.)	1100	2000	150	250	650	4150
pH	10	4.5	7.6	9.7	3.3	
Temp (°C)	29	45	17	20	66	41.5
Pb - Total	72	15	10.5	7.2	90.4	41.3
- Dissolved	32	11	0.7	1.0	77.5	26.0
Zn - Total	2.1	191	2.1	0.74	77.2	104.8
- Dissolved	0.1	1.3	0.07	0.06	0.10	0.67
Cu - Total	0.2	1.5	0.14	0.12	0.12	0.81
- Dissolved	<0.1	196	0.4	<0.4	167.9	120.8
Thiosalts - Total (as S_2O_3)	1220	3122	0	330	3511	2382
- Dissolved (as S_2O_3)	1175	2930	0	310	3395	2274
Thiosulphate - Total	1120	1327	64	380	927	1107
- Dissolved	1117	1251	42	358	890	1060

1. Sampled in September, 1978 (average of Sept. 15, Sept. 18 data).

2. Estimated by BMS personnel.

2.2 Study Effluent Parameters

For the purpose of this study, it was assumed that a feed consisting of a tailings decant stream at ambient temperatures and a relatively warm (15 to 30°C) thickener overflow stream will be treated for thiosalt removal. For most of the options evaluated, the decant passes initially through a settling pond for the removal of residual solids, but the thickener overflow is sent directly for thiosalt treatment in order to conserve heat.

Three situations were considered, and all assumed a constant milling rate of 10,000 t/d. The base case considers a flow and thiosalt discharge similar to that at BMS (720 m^3/h, 24,500 kg/d S_2O_3). Secondly, a similar flow, but a 50% thiosalt discharge were evaluated (720 m^3/h, 12,300 kg/d S_2O_3). Finally, a stream with the same quality, but half the flow of the base case was assumed (360 m^3/h, 12,300 kg/d S_2O_3). The composition of the combined feed stream as well as the target values of the final effluent following thiosalt treatment and pH adjustment are indicated in Table VI.

2.3 Process Flowsheet Definition

The thiosalt treatment processes considered in this study are based on methods evaluated as part of CANMET's thiosalt project(1), as shown in Table VII. Following discussions with suppliers and a review of available technologies, the following were excluded from further evaluation:

TABLE VI

TECHNO-ECONOMIC STUDY EFFLUENT PARAMETERS
(mg/L except pH, temperature)

Parameter	Base Case[1]	100% Flow 50% S_2O_3	50% Flow 50% S_2O_3	Final[2] Effluent
Total thiosalts	1,400	700	1,400	$\leq$ 100
Thiosulphate	700	350	700	$\leq$ 50
Tri-tetra-thionate	700	350	700	$\leq$ 50
Fe	10	10	10	1
Zn	10	10	10	0.5
Pb	5 - 10	5 - 10	5 - 10	0.2
Cu	0.5	0.5	0.5	0.3
Ca	600	600	600	800
SO_4	2,000	2,000	2,000	3,030 - 4,220
Na	500	500	500	500
K and Mg	100	100	100	100
pH	7	7	7	9.5
Temperature	15°C winter 30°C summer	15°C winter 30°C summer	15°C winter 30°C summer	variable with treatment

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1. 720 m^3/h flow, 24,500 kg/d thiosalts.
2. Quality of final effluent expected following thiosalt treatment and pH adjustment.

TABLE VII

SUMMARY OF THIOSALT TREATMENT PROCESSES

[CAUTION: THESE PREVIOUS ESTIMATES ARE NO LONGER VALID]

	Original estimate			
Method	Basis	Annual operating	Capital	Comments
1. Catalyzed air oxidation	1500 U.S. gal/min 5330 mg/L S_2O_3 96 000 lb/d S_2O_3 1st Q, 1981	$8.25 million	$17.5 million	Does not include pre- and post-treatment, test only with synthetic solutions
2. Hydrogen peroxide	77 000 lb/d S_2O_3 4th Q, 1981	$9.02 million	NA	Does not include pre- or post-treatment
3. Chlorine	77 000 lb/d S_2O_3 2nd Q, 1979	$6.0 million	NA	Does not include pre- or post-treatment
4. Ozone	2 million imp. gal/d 1000 mg/L S_2O_3 17 000 lb/d S_2O_3 1975	$1.4 million	NA	Does not include pre- or post-treatment
5. SO_2-Air oxidation	77 000 lb/d S_2O_3 2nd Q, 1982	$6.6 million	NA	Does not include costs for copper catalysts
6. Alkaline oxidation	2000 U.S. gal/min 2000 mg/L S_2O_3 48 000 lb/d S_2O_3 2nd Q, 1982	$5.2 million	$8.9 million	Includes all neutralization costs
7. Ultraviolet oxidation	-	NA	NA	-
8. Reverse osmosis	4000 U.S. gal/min 1500 mg/L S_2O_3 72 000 lb/d S_2O_3 4th Q, 1978	$1 to $1.6 million	$4.3 million	Does not include costs for calcium removal and disposal of concentrated solution
9. Electrodialysis	4000 U.S. gal/min 1500 mg/L S_2O_3 72 000 lb/d S_2O_3 4th Q, 1978	$0.7 to $1.5 million	$4.0 million	Does not include costs for calcium removal and disposal of concentrated solution
10. Electrooxidation	50 000 lb/d S_2O_3 Power = 3.5¢/kWh 2nd Q, 1979	$4.9 million		Does not include pre- or post treatment
11. Reduction by iron	3000 U.S. gal/min 33 000 lb S_2O_3/d 1st Q, 1982	$7.1 million	$38.4 million	Does no include costs for disposal of byproducts of H_2S and FeS
12. Activated carbon treatment	-	NA	NA	-
13. Biological oxidation - RBC	2.3 million gal/d 1250 mg/L S_2O_3 24 000 lb/d S_2O_3 2nd Q, 1980	$430 000	$1.67 million	Does not include pre- and post-treatment
- Noranda Packed ponds		NA	NA	
14. Disposal to the sea	75 cm OD pipeline, 29 km long 109 000 lb/d S_2O_3 1st Q, 1977	$50 000	$5.5 million	Does not include pre-treatment costs

- Ultraviolet oxidation: current equipment is mainly used for disinfection purposes at retention times of ~5 s. High power lamps needed for chemical oxidation are not currently marketed. Also, long retention times required for thiosalt oxidation (>30 min) would make the process exceedingly capital intensive for the high flows considered.
- Electrodialysis: the process has not been evaluated on thiosalt solutions. Discussions with suppliers have indicated that membranes used for sulphate removal may be applicable; however, membrane development work would probably be necessary.

Natural degradation of thiosalts, not considered in the initial review(1) was also investigated in this study.

Process flowsheets for the project were based on data from previous laboratory and pilot-scale thiosalt studies conducted by CANMET, Environment Canada, Noranda Inc., and the University of Waterloo. In some cases (e.g. chemical oxidation using chlorine, peroxide and ozone), data for the treatment of actual mill effluents were unavailable, and additional laboratory studies were undertaken by the Noranda Research Centre and CANMET's Mineral Sciences Laboratories to define the necessary process parameters (e.g. reagent requirements, retention time, etc.).

2.4 Capital Cost Estimates Approach

Capital cost estimates were derived by assuming the thiosalt treatment facility to be a battery limits add-on installation (i.e. construction of a facility on an already prepared and serviced site) to handle the milling effluents of an existing mine/mill complex. The costs considered and excluded for the purpose of this study are summarized in Table VIII. It is assumed that the existing utility main lines (electricity, water) and all service facilities would be adequate to handle the increased load. Steam requirements, where necessary, would however be provided by a new steam plant, and the related capital charges are included in the estimates.

For costing purposes, the process flowsheets were subdivided into process modules (e.g. settling basins, thiosalt oxidation, lime neutralization). Cost estimates of equipment in each module were derived using two methods. In most cases, individual equipment was sized and costed using either suppliers' quotations or recent literature, and the sixth-tenths factor rule(2) was used to generate costs at other than given equipment ratings. In other cases, correlations relating capital cost to process throughput were utilized. All equipment costs are based on mid-1984 values, with the Marshall and Swift index used where necessary for updating earlier costs. The resulting capital charge for the module is the sum of individual process unit costs.

The format used for deriving the overall capital costs for the complete thiosalt treatment facilities are indicated in Table IX. As shown, the bare equipment cost (Item No.1) is the

TABLE VIII

SUMMARY OF BATTERY LIMITS COST COMPONENTS

Includes the cost of	Excludes the cost of
- all process equipment.	- land, site preparation, fencing, access roads and paving, loading and unloading facilities.
- all material and labour to connect and make operable this process equipment, including control rooms, buildings within the battery limits.	- off-site utilities and piping to/from the battery limits from such services.
- indirect field expense, engineering, construction supervision fees.	- support facilities such as the administration building, laboratory, cafeteria, gate house, maintenance shops, fire fighting equipment, communications systems and safety equipment.

TABLE IX

FORMAT FOR CAPITAL COST CALCULATIONS

No.	Item	Base Case
	Primary Module (Thiosalt Destruction)	$'000
1.	Bare Equipment Items	239.00
2.	Misc. Unspecified Items (15% of 1)	35.85
3.	Subtotal Bare Equipment	274.85
4.	Installed Cost (Bare Equipment x 1.45)	398.53
5.	Other Items - Installed Cost	0.00
6.	Subtotal Installed Equipment	398.53
7.	Services: Electric 15%, Instr. 10%, Serv. Piping 10%, Other Serv. 5%	159.41
8.	Building 3000.00 3000.00 2500.00 (sq. ft. at $70/sq. ft.)	210.00
9.	Total Primary Module Cost (A)	767.95
10.	Settling Basin Cost (B)	150.00
11.	Lime Module Cost (C)	8900.00
12.	Total Cost of Modules (A+B+C)	9817.95
13.	Engineering & Construction (25% of Modules)	2454.49
14.	Subtotal	12272.43
15.	Contingency (40% of Subtotal)	4908.97
16.	OVERALL TOTAL CAPITAL COST	17181.40

sum of all process units in the primary (thiosalt destruction) module. A contingency of 15% for miscellaneous equipment was then applied, and the total was factored by 1.45 to obtain the installed cost (Item 4). Other process units where the derived costs includes installation charges were then added (Item 5). Subsequently, factors for services and building costs were considered. The sum of Items 1 through 8 gives the total primary (thiosalt destruction) module. Settling basin and lime neutralization modules were costed separately, and the derived cost was then indicated (Items 10,11). An allowance for engineering and construction supervision expenses (Item 13) was added to the total cost of the modules, and finally, a contingency cost was considered (Item 15) to generate the overall total capital cost (Item 16).

2.5 Basis for Operating and Maintenance Costs

The following parameters contributing to operating and maintenance costs were considered:

- reagents (lime, chlorine, etc.);
- utilities (electricity, steam, fuel);
- operating man-power;
- replacement of components (e.g. membranes).

For each alternative, reagent and utility requirements were based on process conditions. Mid-1984 reagent unit costs are considered, assuming delivery to northeastern New Brunswick.

Electricity rates quoted are those paid by BMS during 1984. Costs for maintenance, taxes, and insurance were taken as 4.5% of the overall total capital costs. The format used to calculate the total operating costs is indicated in Table X. Although by-products may be generated in some of the schemes (e.g. sulphur), no credits were considered against operating costs.

TABLE X

FORMAT FOR OPERATING COST CALCULATIONS

Operating Cost	'000/y.
Materials, Supplies, Utilities & Labour	13769.00
Lime	1664.00
Maint., Taxes & Insurance (4.5% of Tot. Cap.)	773.16
TOTAL OPERATING COST	16206.16

2.6 Annualized Capital Charges

In order to compare with different capital and operating costs, it is desirable to calculate an annualized capital charge, to be added to the annual operating cost to provide an overall "figure of merit".

Annualized capital charges are based on the concept of opportunity cost, i.e. the earnings required to achieve the company's minimum hurdle rate on that capital. The minimum hurdle rate, for risk-free investment, is a function of the company's average cost of capital and is estimated to be 15%, expressed as after-tax discounted cash flow (DCF) return on investment. The effect of taxation must be taken into account, since a capital expenditure provides a "tax shield" spread over a number of future years, whereas an annual cost provides an immediate tax shield. Detailed tax calculations were carried out for hypothetical but realistic cases, from which it is estimated that the net present value of the tax shield is typically 40% of the capital investment, taking into account capital cost allowance, earned depletion and investment tax credit. The annual after-tax earnings required to provide a 15% DCF return on a \$100 capital investment is therefore calculated as follows:

$$(100-40)\ \frac{i\ (1 + i)^n}{(1 + i)^n-1} = \$9.59$$

Where i = discount rate = 15%
n = project life = 20 years

The pretax earnings, based on the effective tax rate of 52% would be:

$$\frac{\$9.59}{(1-.52)} = \$19.98$$, which is rounded to 20% of capital

Thus, annualized capital charges are taken as 20% of capital cost.

3.0 THIOSALT TREATMENT FLOWSHEETS AND COSTS

3.1 Air Oxidation

Air oxidation alone has not been found suitable for the elimination of thiosalts. When aerated under sterile laboratory conditions at pH 7 for up to 4 months, solutions of thiosulphate and polythionates have shown minimal changes in thiosalt concentrations([1]). Biological and catalyzed air oxidation tests have however shown more successful results.

3.1.1 Biological Air Oxidation

Bacteria have been shown capable of oxidizing thiosalts in a number of studies, as shown in Table XI. Pilot tests using an aerated rock-filled pond at Brunswick Mining were conducted between 1979 and 1984, and the highlights of these studies are presented in Appendix A.

Process Flowsheet

Based on the results of the BMS pilot tests, the following process flowsheet is envisaged (Figure 3):

Thiosalt-containing mill tailings decant is sent to one of two settling ponds for settleable solids removal. Subsequently, this stream and warm thickener overflows from the mill pass to an earthen four-cell Aerated Rock Pond (ARP), operated at a retention time of 3 days (based on void volume). Air is distributed throughout the pond by a piping grid laid out on the pond bottom. The rocks (5 to 10 cm dia.) provide surfaces for bacterial attachment, and also act as a diffusing medium for the air. During efficient treatment, the pH of the pond water is

TABLE XI

SUMMARY OF BIOLOGICAL AIR OXIDATION STUDIES

Reference	Type of Reactor	Feed	Process Conditions
Schmidt and Conn, 1971(3)	Lab-scale Activated Sludge	1000 mg/L, S_2O_3, Synthetic	Retention - 18 h, Room Temp.
B.C. Research, 1973(4)	Lab-scale Rotating Biological Contactor (RBC)	Synthetic and BMS Mill Effluent	8.5 and 20°C, 70 to 240 L/day (Ret. = 40 to 130 min.)
Environment Canada, 1980(5)	1. Lab-scale RBC 2. Pilot-scale RBC	Synthetic and BMS Mill Effluents	Lab - 2 to 30°C, 80-400 L/d Pilot - 30 to 35°C, 5.5 to 10.9 m^3/d
Meuller, 1977(6)	1. Lab-scale Packed Column 2. Pilot-scale Packed Column	Synthetic and Meggen Mill Effluents	Lab - 7 L Vol. Pilot - 250 L Column, 6.75 m^3 Tower Scale
Noranda, 1979(7)	1. Lab-scale Mixed Reactors) 2. Lab-scale Packed Reactors)	Synthetic BMS Mill Effluents	Two in Series, 4 L each 3-29°C, 4-33 h Retention 22.4 L Volume, Ambient Temp., 2-3 h Retention.
Noranda, 1979 to 1984 (Results in Appendix A)	3. Pilot Scale Aerated Ponds) 4. Pilot Scale Aerated Rock Ponds)	BMS Mill Effluents	250 m^3 Volume Ambient Temp. 125 m^3 Volume Ambient Temp.

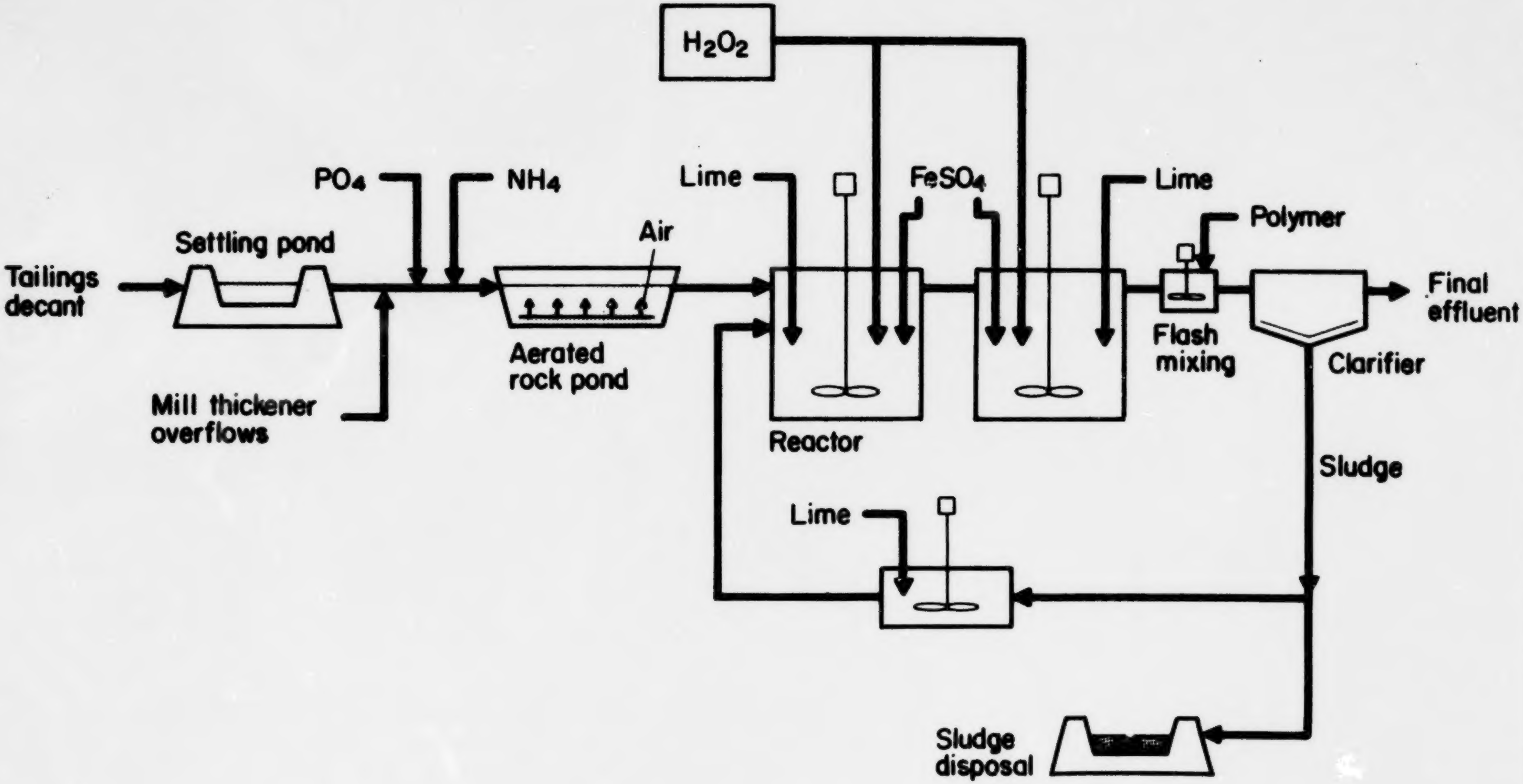

Figure 3 - Biological Air Oxidation

usually below 2.5 (with actual levels dependent on the thiosalt concentrations in the feed).

One settling pond is in use, while the second is being emptied of solids. The four cells of the Aerated Rock Pond are operated in parallel, and any one can be taken out of service as needed for maintenance. There is no provision for continual addition of heat during winter months as the combined tailings and thickener overflow streams are expected to have a temperature > 5°C during winter, which in pilot tests was sufficient for efficient operation. Insulating floating blankets would however be used to minimize heat losses from the pond surfaces.

To maintain optimal conditions for bacterial growth, phosphate and ammonium ion nutrients are added to the ARP feed at a $S_2O_3:NH_4:PO_4$ ratio of 1000:8:5. Additionally, a chemical trim system is provided using H_2O_2 to supplement biological oxidation during possible upsets. Effluents from the aerated cells flow to neutralization reactors where lime is added to raise the pH to 9. Subsequently, precipitated solids are removed in a clarifier, and the clean overflow is discharged to the receiving waters. The principal process design parameters used for sizing the treatment facilities are summarized in Table XII.

TABLE XII
BIOLOGICAL AIR OXIDATION - PROCESS DESIGN PARAMETERS

Parameter	Design Basis		
	720 m^3/h. 24,500 kg/d S_2O_3	720 m^3/h. 12,300 kg/d S_2O_3	360 m^3/h. 12,300 kg/d S_2O_3
Retention time (days)	3	1.5	3
Total pond volume (at 50% void fraction, m^3)	52,000	26,000	26,000
Total pond surface area (m^2)	34,000	18,000	18,000
Air input basis, m^3/min. m^3 liquid volume	0.02	0.02	0.02
Air input (m^3/min.)	520	260	260
Nutrient addition, S_2O_3:NH_4:PO_4	1000:8:5	1000:8:5	1000:8:5
Diammonium Phosphate kg/d	168	84	84
Ammonium Sulphate, kg/d	600	300	300

Cost Estimates

Capital and operating cost estimates for the biological air oxidation of thiosalts are presented in Table XIII for the three cases considered. The total cost of the facilities is estimated to range between \$9.3 and \$13.5 million, including the cost of pre-treatment (settling ponds) and lime neutralization modules. Annual operating costs (excluding amortized capital) range from \$1.5 to \$2.5 million. If chemical trim facilities are deemed necessary to ensure the complete destruction of thiosalts, hydrogen peroxide oxidation would be the most cost-effective (based on the evaluation of chemical oxidation alternatives in Section 3.2). Overall capital costs with H_2O_2 treatment would increase to \$10.9 to \$15.6 million, with annual operating costs of between \$2.2 and \$3.9 million, for the three cases (Table XIV).

TABLE XIII - BIOLOGICAL AIR OXIDATION - COST ESTIMATES

	Base Case	100% Flow 50% Thio	50% Flow 50% Thio
Primary Module	$'000	$'000	$'000
Bare Equipment Items	529.00	295.00	295.00
Misc. Unspectified Items (15% of 1)	79.35	44.25	44.25
Subtotal Bare Equipment	608.35	339.25	339.25
Installed Cost (Bare Equipment x 1.45)	882.11	491.91	491.91
Other Items - Installed Cost	0.00	0.00	0.00
Subtotal Installed Equipment	882.11	491.91	491.91
Services: Electrical 15%, Instr. 10%, Serv. Piping 10%, Other Serv. 5%	352.84	196.77	196.77
Building 800.00 700.00 700.00	56.00	49.00	49.00
Earthworks & Concrete	1,642.00	1,130.00	1,130.00
Total Primary Module Cost (A)	2,932.95	1,867.68	1,867.68
Settling Basin Cost (B)	150.00	150.00	100.00
Lime Module Cost (C)	3,580.00	3,240.00	2,850.00
Total Cost of Modules (A + B + C)	6,662.95	5,257.68	4,817.68
Engineering & Construction (25% of A+B+C)	1,665.74	1,314.42	1,204.42
Subtotal	8,328.69	6,572.10	6,022.10
Contingency (40% of Subtotal)	3,331.48	2,628.84	2,408.84
Washed Rock (Includes 20% Contingency)	1,800.00	900.00	900.00
OVERALL TOTAL CAPITAL COST	13,460.16	10,100.94	9,330.94
Operating Cost	'000/y	'000/y	'000/y
Materials, Supplies, Utilities & Labour	918.00	559.00	559.00
Lime	1,000.00	500.00	500.00
Maint., Taxes & Insurance (4.5% of Tot.Cap.)	605.71	454.54	419.89
TOTAL OPERATING COST	2,523.71	1,513.54	1,478.89

TABLE XIV - BIOLOGICAL AIR OXIDATION WITH POST-TREATMENT - COST ESTIMATES

	Base Case	100% Flow 50% Thio	50% Flow 50% Thio
Primary Module	$'000	$'000	$'000
Bare Equipment Items	879.00	535.00	535.00
Misc. Unspectified Items (15% of 1)	131.85	80.25	80.25
Subtotal Bare Equipment	1,010.85	615.25	615.25
Installed Cost (Bare Equipment x 1.45)	1,465.73	892.11	892.11
Other Items - Installed Cost	0.00	0.00	0.00
Subtotal Installed Equipment	1,465.73	892.11	892.11
Services: Electrical 15%, Instr. 10%, Serv. Piping 10%, Other Serv. 5%	586.29	356.85	356.85
Building 6,500.00 5,500.00 5,500.00	455.00	385.00	385.00
Earthworks & Concrete	1,642.00	1,130.00	1,130.00
Total Primary Module Cost (A)	4,149.03	2,763.96	2,763.96
Settling Basin Cost (B)	150.00	150.00	100.00
Lime Module Cost (C)	3,580.00	3,240.00	2,850.00
Total Cost of Modules (A + B + C)	7,879.03	6,153.96	5,713.96
Engineering & Construction (25% of A+B+C)	1,969.76	1,538.49	1,428.49
Subtotal	9,848.78	7,692.45	7,142.45
Contingency (40% of Subtotal)	3,939.51	3,076.98	2,856.98
Washed Rock (Includes 20% Contingency)	1,800.00	900.00	900.00
OVERALL TOTAL CAPITAL COST	15,588.29	11,669.43	10,899.43
Operating Cost	'000/y	'000/y	'000/y
Materials, Supplies, Utilities & Labour	2,244.00	1,222.00	1,222.00
Lime	1,000.00	500.00	500.00
Maint., Taxes & Insurance (4.5% of Tot.Cap.)	701.47	525.12	490.47
TOTAL OPERATING COST	3,945.47	2,247.12	2,212.47

A review of the cost components (Tables BI, BII - Appendix B) indicates the major capital items to include the rock packing, earthworks, blowers, and piping. Operating expenses are mainly for electrical power, lime, and nutrients in biological oxidation alone, and for reagents (hydrogen peroxide, ferrous sulphate) when chemical trim is considered.

The total annual costs (operating and annualized capital costs) are summarized in Table XV. For biological treatment alone, annual costs range from \$3.4 to \$5.2 million. This translates into \$0.91 to \$1.43/t ore milled, and water treatment costs of \$0.56 to \$1.06/m^3. For biological oxidation with a hydrogen peroxide trim, total annual costs fall in the range of \$4.4 to \$7.1 million, resulting in \$1.20 to \$1.94/t ore milled, and water treatment costs of \$0.72 to \$1.39/m^3.

TABLE XV

BIOLOGICAL AIR OXIDATION COST SUMMARIES

Item	Base Case[1.]	100% Flow 50% S_2O_3	50% Flow 50% S_2O_3
Biological Air Oxidation			
Annualized total costs (millions)	\$ 5.22	\$ 3.53	\$ 3.35
Cost/m^3 water treated	\$ 0.82	\$ 0.56	\$ 1.06
Cost/t ore milled	\$ 1.43	\$ 0.96	\$ 0.91
Biological Air Oxidation with H_2O_2 treatment			
Annualized total cost (millions)	\$ 7.06	\$ 4.58	\$ 4.39
Cost/m^3 water treated	\$ 1.14	\$ 0.72	\$ 1.39
Cost/t ore milled	\$ 1.94	\$ 1.25	\$ 1.20

1. 720 m^3/h, 24,500 kg/d thiosalts.

Comments on the Process

Despite promising laboratory and pilot-scale tests, uncertainties remain regarding the feasibility of a full-scale biological air oxidation facility:

- The pilot plant at BMS has treated successfully a maximum flow of only 1.2 to 1.8 m^3/h. A full-scale facility would treat a flow 600 times larger - a significant scaling-up factor.

- Heavy metals (especially lead) have been shown to interfere with biological activity. Although lead levels in the combined mill effluents are expected to be generally acceptable, occasionally excessive concentrations may disrupt the biological oxidation process.

- The biological process required extensive start-up times (one to two months). Interruptions in the process (e.g. strikes, mill maintenance down-time) would require lengthy restarting. During this time, unacceptable effluents would be released, unless full-scale chemical oxidation is provided.

- Unlike organic decomposition, the destruction of thiosalts by bacteria does not result in the generation of voluminous biological solids. Some solids may form (e.g. sulphur), but insufficient information is available to determine potential detrimental effects over a pond life of 15 to 25 years.

- During winter operation, freezing problems have interfered with pilot plant operations. Such problems may create even more difficulties in the operation of a full-scale plant.

3.1.2 Catalyzed Air Oxidation

Research studies have been conducted at the University of Waterloo aimed at the development of catalytic oxidation of thiosalts. This work has resulted in two main methods using the following catalytic agents:

- polyvinyl pyridine (PVP) with bound copper([8])
- intermediate products from the copper smelting process (e.g. reverberatory matte (30% Cu), converter white metal (70% Cu))([9]).

Both of the methods are reviewed in the following sections.

3.1.2.1 PVP Copper Catalyzed Air Oxidation

Poly-4-Vinyl Pyridine is a granular, insoluble, cross-linked polymer derived from 4-Vinyl Pyridine. Its typical properties are presented in Table XVI. For the thiosalt oxidation studies, a copper complex was prepared by reacting the polymer with a methanolic solution of $CuSO_4$. Based on the results of the University of Waterloo laboratory studies, the following process flowsheet (Figure 4) is envisaged:

Thiosalt-containing mill tailings decant is pumped to one of two settling ponds for settleable solids removal. Subsequently, this stream, and warm thickener overflows from the mill are air-oxidized in two equally-sized aerated earthen basins containing a 2% PVP-Cu resin slurry and operated in parallel at a retention time of 16 hours. [It is assumed that one-stage treat-

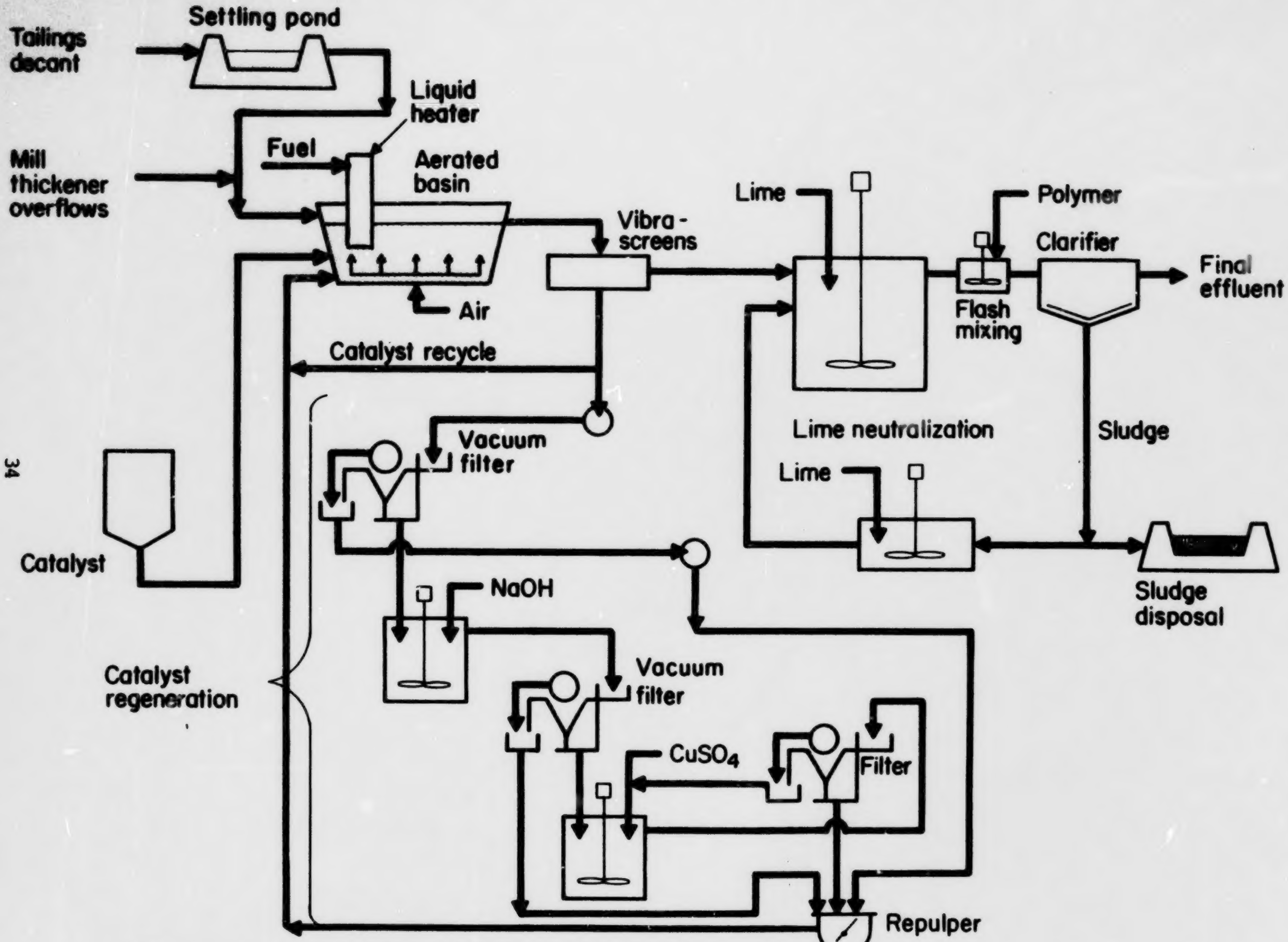

Figure 4 - PVP-Cu Catalyzed Air Oxidation

TABLE XVI

POLYVINYL PYRIDINE - TYPICAL PROPERTIES

Appearance	Off-white to tan granular solid
Bulk density, g/mL	0.45
Skeletal Density, g/cm^3	1.154
Moisture, % by weight	
dry	0.5% - 2.0%
wet	36 - 39%
Apparent pKa	5.8
Hydrogen ion capacity, meq/g based on dry polymer	
in water	8.8
in toluene	8.5
Percentage swelling from free base to hydrochloride form, in water	100%
Percentage swelling from dry state to solvent-saturated state	
methanol	70 - 75%
acetone	30 - 35%
water	33 - 37%
isopropanol	13 - 17%
toluene	8 - 12%
ethyl acetate	3 - 6%
hexane	0%
Temperature stability	>100°C

ment will be achievable although in laboratory tests, two-stage oxidation using two different resin materials proved more effective.] Heat is added as needed, using submerged combustion, to maintain water temperature in the ponds at 30°C. Static mixers are utilized for air addition. During efficient treatment, the pH of the pond water would be below 2.5 due to conversion of thiosalts to SO_4 although actual pH levels are dependent on thiosalt concentrations in the feed.

One of the settling ponds would be in use continuously, while the other is emptied of solids. The two aerated cells would be operated simultaneously and any one can taken out of service, as needed, for maintenance. Floating insulation blankets on the pond surfaces will minimize heat losses.

Discharge from the aerated basins is screened, with the treated water sent for lime neutralization, and 90% of the PVP-Cu resin in the underflow recycled to the process. Since the resin becomes deactivated due to loss of copper from the polymer matrix, as well as adsorption of thiosalts, the remaining 10% of the resin will be sent continuously for regeneration, consisting of:

- dewatering to 40% solids using a filter press;
- contacting with caustic to remove adsorbed thiosalts;
- washing and filtering;
- contacting with a solution of copper sulphate;
- washing and filtering to remove excess $CuSO_4$;
- reslurring of the regenerated resin and recycle to the process.

The principal process design parameters used for sizing the catalyzed air oxidation treatment facilities are summarized in Table XVII.

TABLE XVII

PVP-Cu CATALYZED AIR OXIDATION - PROCESS DESIGN PARAMETERS

Parameter	Design Basis		
	720 m^3/h. 24,500 kg/d S_2O_3	720 m^3/h. 12,300 kg/d S_2O_3	360 m^3/h. 12,300 kg/d S_2O_3
Retention time (days)	0.66	0.66	0.66
No. of basins	2	2	2
Total pond volume (m^3)	11,500	11,500	6,000
Total pond surface area (m^3)	5,400	5,400	2,800
Air Input basis (m^3/min.m^3 liquid volume)	0.08	0.08	0.08
Air input (m^3/min)	920	920	480
Operating temp.	30°C	30°C	30°C
Catalyst conc. (% in pond liquid)	2	2	2
Catalyst use before regeneration (days)	10	20	10
Catalyst life (yrs)	1	2	1

Cost Estimates

Capital and operating cost estimates for the PVP-Cu Catalyzed Air Oxidation process are presented in Table XVIII for the three cases considered. Overall capital costs range from \$19.3 to \$32.2 million, with a major portion being due to the initial resin catalyst charge (\$9 million to \$18 million). Annual operating costs fall between \$10.1 and \$19.6 million. A review of the cost components (Table B3 - Appendix B) indicates other major capital expenditures to include the aeration system (blowers, piping, static mixers) and the catalyst regeneration system. Major operating expenses include resin catalyst replace-

TABLE XVIII - PVP-Cu CATALYZED AIR OXIDATION
COST ESTIMATES

	Base Case	100% Flow 50% Thio	50% Flow 50% Thio
Primary Module	$'000	$'000	$'000
Bare Equipment Items	1,675.30	1,137.00	1,137.00
Misc. Unspectified Items (15% of 1)	251.30	170.55	170.55
Subtotal Bare Equipment	1,926.30	1,307.55	1,307.55
Installed Cost (Bare Equipment x 1.45)	2,793.56	1,895.95	1,895.95
Other Items - Installed Cost	0.00	0.00	0.00
Subtotal Installed Equipment	2,793.56	1,895.95	1,895.95
Services: Electrical 15%, Instr. 10%, Serv. Piping 10%, Other Serv. 5%	1,117.43	758.38	758.38
Building 3,250.00 2,500.00 2,500.00	227.50	175.00	175.00
Earthworks & Concrete	254.10	185.00	160.00
Total Primary Module Cost (A)	4,392.59	3,014.33	2,989.33
Settling Basin Cost (B)	150.00	150.00	100.00
Lime Module Cost (C)	3,580.00	3,240.00	2,850.00
Total Cost of Modules (A + B + C)	8,122.59	6,404.33	5,939.33
Engineering & Construction (25% of A+B+C)	2,030.65	1,601.08	1,484.83
Subtotal	10,153.23	8,005.41	7.424.16
Contingency (40% of Subtotal)	4,061.29	3,202.16	2,969.66
Initial Catalyst Charge (Incl. 20% Conting.)	18,000.00	9,000.00	9,000.00
OVERALL TOTAL CAPITAL COST	32,214.53	20,207.57	19,393.82
Operating Cost	'000/y	'000/y	'000/y
Materials, Supplies, Utilities & Labour	17,124.00	9,162.00	8,762.00
Lime	1,000.00	500.00	500.00
Maint., Taxes & Insurance (4.5% of Tot.Cap.)	1,449.65	909.34	872.72
TOTAL OPERATING COST	19,573.65	10,571.34	10,134.72

ment, fuel for submerged combustion, power, and reagents.

The total annual costs (operating and annualized capital costs) are summarized in Table XIX, with costs ranging from \$14.0 to \$26.0 million. This translates into \$3.84 to \$7.14/t ore milled, or water treatment costs of \$2.31 to \$4.42/m^3.

TABLE XIX

PVP-Cu CATALYZED AIR OXIDATION COST SUMMARIES

Item	Base[1.] Case	100% Flow 50% S_2O_3	50% Flow 50% S_2O_3
Annualized total cost (millions)	\$26.02	\$14.61	\$14.01
Cost/m^3 water treated	\$ 4.12	\$ 2.31	\$ 4.42
Cost/t ore milled	\$ 7.14	\$ 4.00	\$ 3.84

1. 720 m^3/h, 24,500 kg/d thiosalts.

Comments on the Process

The PVP-Cu resin catalyzed air oxidation process has been tested to date only in laboratory-scale studies using synthetic effluents. Although successful oxidation of thiosalts was achieved for feed concentrations of up to 1600 mg/L as S_2O_3, the process is considered unviable for a number of reasons:

- Unsatisfactory resin regeneration - Tests at the University of Waterloo have indicated a significant loss in resin activity (20 to 60%) following initial catalyst regeneration. Although there is insufficient information to indicate resin life, the poor regeneration characteristics suggest it may be quite limited, thus requiring frequent make-up or replacement.

- High Cost of Resin Material - Currently, the resin is supplied principally by Reilly Tar and Chemical Co., Indianapolis, Indiana, at a cost of $50 US/kg. This price is similar to that of ion exchange resins, and it is unlikely that cost can be reduced significantly, even with large volume purchases (e.g. to $2 to $10/kg range).

3.1.2.2 Copper Sulphide Catalyzed Air Oxidation

The flowsheet of the copper sulphide catalyzed air oxidation process (using converter white metal) is illustrated in Figure 5. Thiosalt-containing mill tailings decant is initially pumped to one of two settling ponds for settleable solids removal. The clarified effluent and warm thickener overflows from the mill are then oxidized in two equally-sized aerated earthen basins operated at a retention time of 24 h and containing a 4% white metal slurry. Submerged combustion is used to maintain the pond water temperature of 30°C, and air is introduced through static mixers for thiosalt oxidation requirements and mixing. Based on the results of the University of Waterloo studies, the pond pH is expected to remain in the range of 5 to 7 (in spite of sulphate generation) due to the buffering action of the white metal.

Discharge from the aerated basins is screened, with the treated water sent for lime neutralization, and the white metal solids will be recycled to the process. Approximately 3% of

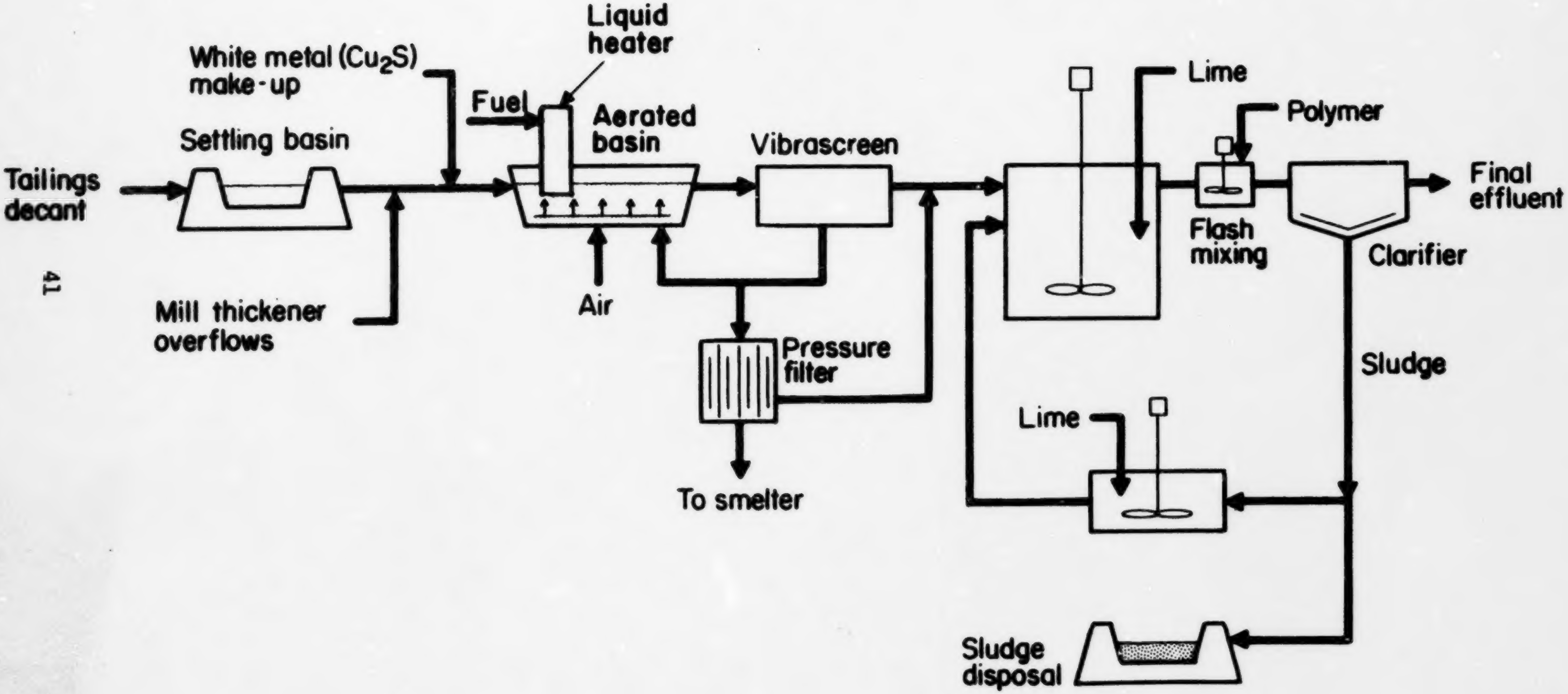

Figure 5 - Copper Sulphide Catalyzed Air Oxidation

the solids will be taken out continuously, pressure filtered, and sent to a copper smelter for reprocessing.

One of the settling ponds is operated continuously, while the second is being emptied of accumulated solids which could be disposed of in the tailings pond. The two aerated cells would be operatud in parallel, and one can be taken out of service for maintenance, etc. as required. To minimize heat losses, floating insulation blankets would cover the pond surfaces.

The principal process design parameters used in sizing the copper catalyzed air oxidation facilities are summarized in Table XX.

TABLE XX

COPPER SULPHIDE CATALYZED AIR OXIDATION - PROCESS DESIGN PARAMETERS

Parameter	Design Basis		
	720 m^3/h. 24,500 kg/d S_2O_3	720 m^3/h. 12,300 kg/d S_2O_3	360 m^3/h. 12,300 kg/d S_2O_3
Retention time (days)	1	1	1
No. of basins	2	2	2
Total pond volume (m^3)	17,500	17,500	9,000
Total pond surface area (m^2)	7,000	7,000	4,700
Air input basis (m^3/min.m^3 liquid volume)	0.09	0.09	0.09
Operating temp (°C)	30	30	30
White metal conc. (% solids)	4	4	4
Catalyst use before regeneration (months)	1	1	1

Cost Estimates

Capital and operating expenses for the Copper Catalyzed Air Oxidation process are summarized in Table XXI. Overall capital costs range from $9.7 to $14.8 million for the three cases considered, and corresponding annual operating costs vary from $2.4 to $4.2 million. An examination of the cost components (Table B4 - Appendix B) indicates major portions of capital related to the pond aeration system (blower, aerators, piping), submerged combustion equipment, and the initial catalyst charge. Operating expenses are made-up principally of white metal reprocessing/transportation costs, fuel for the submerged combustion system, electrical power, and white metal make-up requirements.

The total annual costs (operating and annualized capital costs) are summarized in Table XXII. As shown, total annual costs range from $4.3 to $7.2 million, resulting in charges of $1.19 to $1.96/ton ore milled, or water treatment costs of $0.84 to $1.38/$\mathrm{m}^3$.

TABLE XXII

COPPER CATALYZED AIR OXIDATION - COST SUMMARIES

Item	Base Case[1.]	100% Flow 50% S_2O_3	50% Flow 50% S_2O_3
Annualized total cost (millions)	\$ 7.17	\$ 5.33	\$ 4.34
Cost/m^3 water treated	\$ 1.14	\$ 0.84	\$ 1.38
Cost/t ore milled	\$ 1.96	\$ 1.46	\$ 1.19

1. 720 m^3/h, 24,500 kg/d thiosalts.

TABLE XXI - COPPER CATALYZED AIR OXIDATION - COST ESTIMATES

	Base Case	100% Flow 50% Thio	50% Flow 50% Thio
Primary Module	$'000	$'000	$'000
Bare Equipment Items	1,804.50	1,216.00	1,091.00
Misc. Unspectified Items (15% of 1)	270.68	182.40	163.65
Subtotal Bare Equipment	2,075.18	1,398.40	1,254.65
Installed Cost (Bare Equipment x 1.45)	3,009.00	2,027.68	1,819.24
Other Items - Installed Cost	0.00	0.00	0.00
Subtotal Installed Equipment	3,009.00	2,027.68	1,819.24
Services: Electrical 15%, Instr. 10%, Serv. Piping 10%, Other Serv. 5%	1,203.60	811.07	727.70
Building 3,250.00 2,500.00 2,500.00	227.50	175.00	175.00
Earthworks & Concrete	382.50	293.00	268.00
Total Primary Module Cost (A)	4,822.61	3,306.75	2,989.94
Settling Basin Cost (B)	150.00	150.00	100.00
Lime Module Cost (C)	2,800.00	2,500.00	2,100.00
Total Cost of Modules (A + B + C)	7,772.61	5,956.75	5,189.94
Engineering & Construction (25% of Modules)	1,943.15	1,489.19	1,297.48
Subtotal	9,715.76	7,445.94	6,487.42
Contingency (40% of Subtotal)	3,886.30	2,978.38	2,594.97
Initial Catalyst Charge (incl. 20% Conting.)	1,152.00	576.00	576.00
OVERALL TOTAL CAPITAL COST	14,754.06	11,000.32	9,658.39
Operating Cost	'000/y	'000/y	'000/y
Materials, Supplies, Utilities & Labour	3,480.00	2,595.00	1,9[illegible].00
Lime	75.00	37.50	37.50
Maint., Taxes & Insurance (4.5% of Tot.Cap.)	663.93	495.01	434.63
TOTAL OPERATING COST	4,218.93	3,127.51	2,412.13

Comments on the Process

The copper sulphide catalyzed air oxidation process has been developed and tested in laboratory-scale continuous studies at the University of Waterloo using both synthetic and actual BMS milling effluents. In both cases, successful results were achieved with more efficient treatment taking place when actual solutions were used, perhaps due to the catalytic action of metals (e.g. copper, lead, zinc) present in the mill effluents. The process has proven self-buffering in the laboratory tests, within a pH range of 5-6, despite the generation of SO_4. The phenomenon is attributed to solution and reprecipitation of copper from the white metal catalyst utilized although the results to date have not defined the maximum buffering capacity.

Extensive additional test work is required before this process can be implemented, with a need to define the following:

- White metal requirements as a function of quantity of thiosalt treated.
- Feasibility of continuous treatment under variable feed stream conditions.
- Need, if any, for neutralization.
- Optimal white metal slurry concentration.
- Optimal temperature and retention time.

3.2 Chemical Oxidation

Chemical oxidants, such as chlorine, hydrogen peroxide, and ozone, have all been found to oxidize thiosalts to sulphate(^{1}). Each of the oxidants will be discussed in the following sections.

3.2.1 Hydrogen Peroxide

Hydrogen peroxide is a widely used chemical reagent, quite stable under normal conditions when properly stored. At normal ambient temperatures, there is less than 1% loss of H_2O_2 per year in large containers (e.g. bulk storage tanks) and roughly a 2% per year loss in smaller containers (e.g. drums)(^{11}). The chemical and physical properties of H_2O_2 solutions are presented in Table XXIII.

Process Flowsheet

Bench-scale treatability studies have been conducted at the Noranda Research Centre, and at CANMET's Mineral Sciences Laboratories using actual BMS solutions. The results of these studies suggest the following process flowsheet (Figure 6):

Thiosalt-containing mill tailings decant is sent to one of two settling ponds for settleable solids removal. Subsequently, this stream and warm thickener overflows from the mill are combined and pumped for treatment in three agitated reactors operated in series. For optimal treatment, oxidation takes place at pH 7(^{12}) with hydrogen peroxide and ferrous

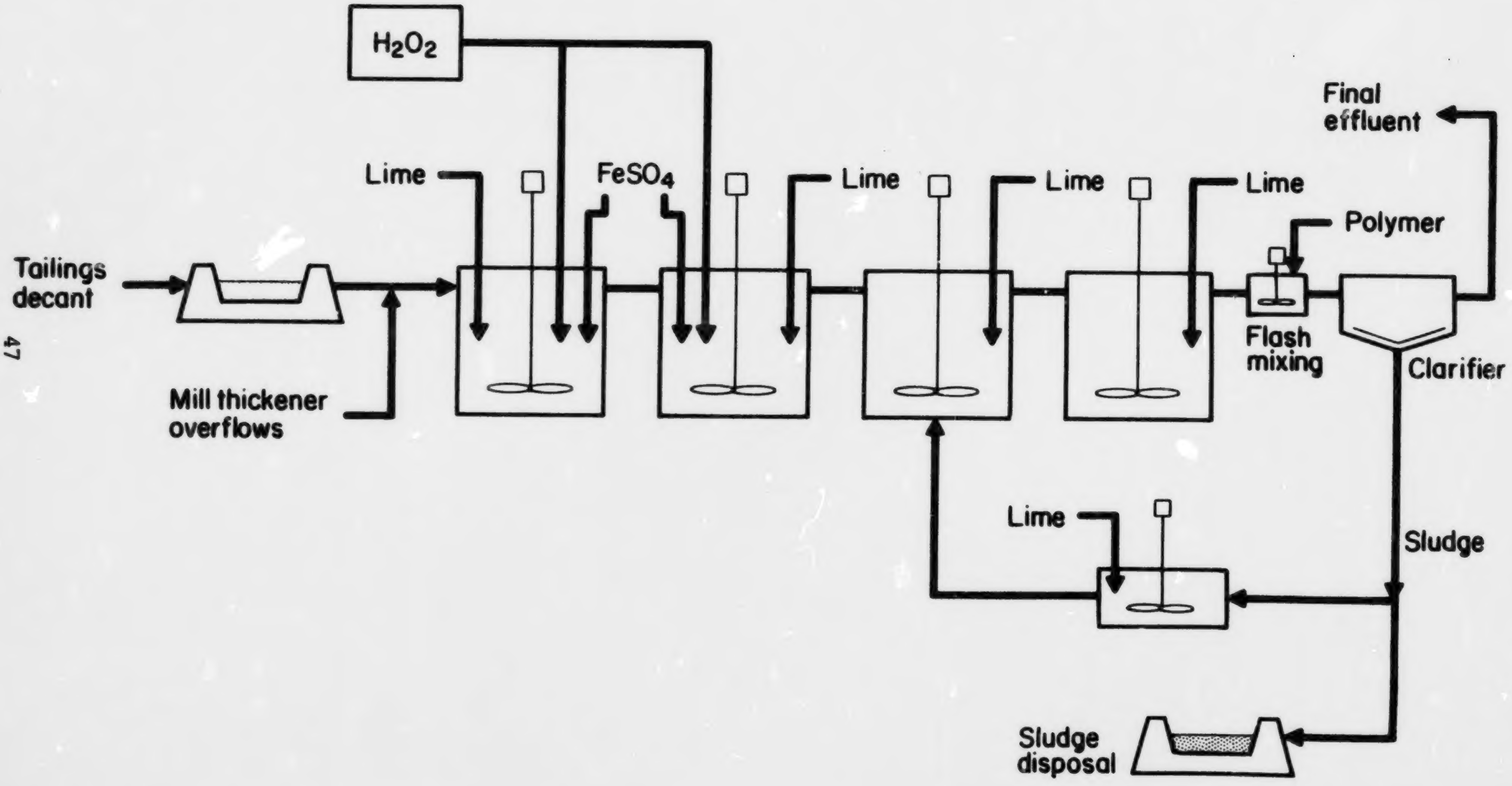

Figure 6 - Hydrogen Peroxide Oxidation

TABLE XXIII - CHEMICAL AND PHYSICAL PROPERTIES OF HYDROGEN PEROXIDE (11)

Concentrations, weight %	**35**	**50**	**70**
"Volume strength" @ 0°C & 1 atm.	130	197	300
Active oxygen content, weight %	16.5	23.5	32.9
H_2O_2, g/L	396	600	903
Specific gravity @ 20°C/4°C	1.13	1.20	1.29
Lbs/gal @ 20°C	9.4	10.0	10.8
Boiling point, °C	108	114	126
Boiling point, °F	226	237	258
Freezing point, °C	- 33	- 52	- 40
Freezing point, °F	- 27	- 62	- 40
Viscosity @ 25°C (centipoise)	1.0	1.06	1.12
Refractive index @ 25°C	1.355	1.366	1.381
Dielectric constant @ 20°C	83	83	82
Total vapor pressure @ 30°C (mm Hg)	23.3	18.3	10.1
Partial pressure @ 30°C (mm Hg)	0.28	0.56	1.17
Heat of dilution, cal/g mole @ 25°C & 1 atm.	- 84	- 178	- 381
Surface tension, dynes/cm @ 20°C	74.6	75.6	77.3

sulphate added to promote the reaction (ferrous sulphate has been found necessary as a catalyst in laboratory studies). Lime is added for pH control, and the reactors are each operated at a 15 min. retention time. Reagent requirements, based on laboratory tests with actual BMS effluents, are 1.15 g H_2O_2/g S_2O_3 and 65 mg/L Fe. Mixing is provided by mechanical agitators.

Following oxidation, the effluent is sent for neutralization to pH 9 and solids/liquid separation in a lime treatment plant. The clean overflow is then discharged to the receiving waters.

The principal design parameters used in sizing the H_2O_2 oxidation facilities are presented in Table XXIV.

TABLE XXIV

H_2O_2 OXIDATION - PROCESS DESIGN PARAMETERS

Parameter	Design Basis		
	720 m^3/h. 24,500 kg/d S_2O_3	720 m^3/h. 12,300 kg/d S_2O_3	360 m^3/h. 12,300 kg/d S_2O_3
H_2O_2: S_2O_3 weight ratio requirements	1.15	1.15	1.15
Fe addition rate (mg/L)	65	65	65
Total reactor retention time (min)	45	45	45
Agitator sizing (kW/m^3)	0.16	0.16	0.16
No. of reactors	3	3	3

Cost Estimates

Table XXV presents the cost estimates for the three cases involving hydrogen peroxide oxidation of thiosalts.

TABLE XXV - H_2O_2 OXIDATION - COST ESTIMATES

	Base Case	100% Flow 50% Thio	50% Flow 50% Thio
Primary Module	$'000	$'000	$'000
Bare Equipment Items	405.00	350.00	295.00
Misc. Unspectified Items (15% of 1)	60.75	52.50	44.25
Subtotal Bare Equipment	465.75	402.50	339.25
Installed Cost (Bare Equipment x 1.45)	675.34	583.63	491.91
Other Items - Installed Cost	0.00	0.00	0.00
Subtotal Installed Equipment	675.34	583.63	491.91
Services: Electrical 15%, Instr. 10%, Serv. Piping 10%, Other Serv. 5%	270.14	233.45	196.77
Building 6,000.00 5,500.00 5,500.00	420.00	385.00	385.00
Total Primary Module Cost (A)	1,365.47	1,202.08	1,073.68
Settling Basin Cost (B)	150.00	150.00	100.00
Lime Module Cost (C)	3,580.00	3,240.00	2,850.00
Total Cost of Modules (A + B + C)	5,095.47	4,592.08	4,023.68
Engineering & Construction (25% of Modules)	1,273.87	1,148.02	1,005.92
Subtotal	6,369.34	5,740.09	5,029.60
Contingency (40% of Subtotal)	2,547.74	2,296.04	2,011.84
OVERALL TOTAL CAPITAL COST	8,917.08	8,036.13	7,041.44
Operating Cost	'000/y	'000/y	'000/y
Materials, Supplies, Utilities & Labour	13,533.00	7,033.00	6,867.00
Lime	1,000.00	500.00	500.00
Maint., Taxes & Insurance (4.5% of Tot.Cap.)	401.27	361.63	316.86
TOTAL OPERATING COST	14,934.27	7,894.63	7,683.86

Overall capital charges are fairly independent of the flow, and range from $7 to $8.9 million. The major part of these costs (~75%) is accounted for by lime neutralization facilities. Annual operating costs total $7.6 to $14.9 million, and are directly related to thiosalt loadings with ~90% accounted for by expenditures for hydrogen peroxide. Table B5 in Appendix B presents itemized cost break-downs of the major capital and operating charges.

The total annual costs (operating and annualized capital costs) are summarized in Table XXVI. As shown, annual costs range from $9.1 to $16.7 million, resulting in charges of $2.49 to $4.58/t ore milled, or water treatment costs of $1.52 to $2.88/m^3.

TABLE XXVI

H_2O_2 OXIDATION COST SUMMARIES

Item	Base[1.] Case	100% Flow 50% S_2O_3	50% Flow 50% S_2O_3
Annualized total cost (millions)	$16.72	$ 9.50	$ 9.09
Cost/m^3 water treated	$ 2.67	$ 1.51	$ 2.88
Cost/t ore milled	$ 4.58	$ 2.61	$ 2.49

1. 720 m^3/h, 24,500 kg/d thiosalts.

Comments on Process

H_2O_2 oxidation is used industrially for the removal of thiosalts from both photographic and steel manufacturing effluents(13). Recently, a gold mine in New Guinea has started using hydrogen peroxide for the destruction of cyanides in its effluents(14). The use of hydrogen peroxide for chemical oxidation of thiosalts in mill effluents offers advantages over other methods, as follows:

- Hydrogen peroxide oxidation technology would be familiar to mill operators (e.g. agitators, mix-tank reactor vessels, etc.). A full-scale facility, based on currently available data could be installed with a high degree of certainty of successful operation.
- The process does not result in by-product contaminants and (unlike, for example, chlorination) the only products of hydrogen peroxide oxidation of thiosalts are SO_4 and H_2O.
- The process is relatively inexpensive in capital because of the simple flowsheet.

3.2.2 Ozone Oxidation

Ozone is a polycondensed form of oxygen. Under normal temperature and pressure conditions, it is an unstable gas which decomposes easily into molecular oxygen and activated oxygen of considerable oxidizing power. The direct oxidation reactions of ozone due to the activated oxygen are usually rapid as ozone has a high redox potential in comparison with the halogens (Table XXVII). Ozone also produces indirect oxidation reactions as part of the ozone dissolved in water decomposes to form free radicals $(OH^{\bullet})$ which react very quickly to oxidize organic and inorganic

TABLE XXVII

REDOX POTENTIALS OF OZONE AND THE HALOGENS(17)

Parameter	Redox Potential (volts vs hydrogen)
	E.
$O_3 + 2H^+ + 2e^- = O_2 + H_2O$	2.07
$HOCl + H^+ + e^- = Cl^- + H_2O$	1.49
$Cl_2 + 2e^- = 2Cl^-$	1.36
$HOBr + H^+ + 2e^- = Br^- + H_2O$	1.33
$O_3 + H_2O + 2e^- = O_2 + 2OH^-$	1.24
$ClO_2 + e^- = ClO^-_2$	1.15
$Br_2 + 2e^- = 2Br^-$	1.07
$ClO_2\ aq + e^- = ClO^-_2$	0.95
$OCl^- + H_2O + 2e^- = Cl^- + 2OH^-$	0.90
$OBr^- + H_2O + 2e^- = Br^- + 2OH^-$	0.70
$I_2 + 2e^- = 2I^-$	0.54
$I_3^- + 2e^- = 3I^-$	0.53
$OI^- + H_2O + 2e^- = I + 2OH^-$	0.49

compounds present in the water(15). Ozone decomposition is accelerated by alkaline conditions which favour the formation of the OH free radicals.

Ozone can be produced from oxygen in the air or from pure oxygen. The selection of air or oxygen as the feed gas affects the performance of the ozonator, as with the same power consumption and gas flow, oxygen produces about twice as much ozone as does air. The use of oxygen for ozone generation is generally more cost-effective for ozone requirements greater than 500 kg/d. Because of its unstable form, ozone is generated on site and is used immediately. For continuous operation, the feed gas is circulated between two electrodes separated by a narrow gap. The application of a high voltage alternating current creates an electric field throughout the gap which acts on the oxygen to form ozone. Ozone generators are of two main types:

- ozonators with tubular electrodes
- ozonators with flat electrodes

Ozone is widely used for the treatment of potable water, expecially in France, Switzerland, and Germany(15). A recent U.S. EPA sponsored study(17) has reviewed industrial water and wastewater treatment with ozone, and has indicated uses in such industries as photo processing, electroplating, and petroleum refineries.

Ozone oxidation of thiosalts in mine effluents was studied initially by Environment Canada's Wastewater Technology Centre(1). Based on those studies, the following reactions were proposed:

$$2S_2O_3^{2-} + 5O_3 + 2H_2O \rightarrow 4SO_4 + 3\ 1/2\ O_2 + 4H^+$$

$$5S_2O_3^{2-} + 2O_3 + 7O_2 + 5H_2O \rightarrow 10SO_4^{2-} + 10H^+$$

More recently, laboratory ozonation studies were conducted by CANMET's Mineral Sciences Laboratories using synthetic and actual BMS mill effluents([18]). The results indicated that ozonation of thiosalts proceeded very rapidly, with the reaction mechanism affected by the pH, solution composition, ozone addition rate, and oxygen concentration. At low pH, H_2S and elemental sulphur by- products were formed resulting in decreased O_3 requirements. Reagent requirements were observed to range between 1.6/1 to 4/1 O_3/S_2O_3 molar ratios.

Process Flowsheet

A flowsheet for the treatment of thiosalt containing milling effluents by ozonation is presented in Figure 7. Thiosalt-containing mill tailings decant is pumped to one of two settling ponds for settleable solids removal. Subsequently, this stream and warm thickener overflows from the mill are sent for treatment to three equally-sized covered ozone reactors operated in series, with a total retention time of 60 minutes. Piping is arranged to enable any one reactor to be taken out of service, if needed, and maintain efficient operation with the remaining two reactors. Agitation in the reactors is provided by mechanical mixers, and the liquid in the reactors is maintained at pH ~9 with lime. Following ozonation, the effluent is sent for solid/liquid separation, and the clean overflow is discharged to the receiving waters.

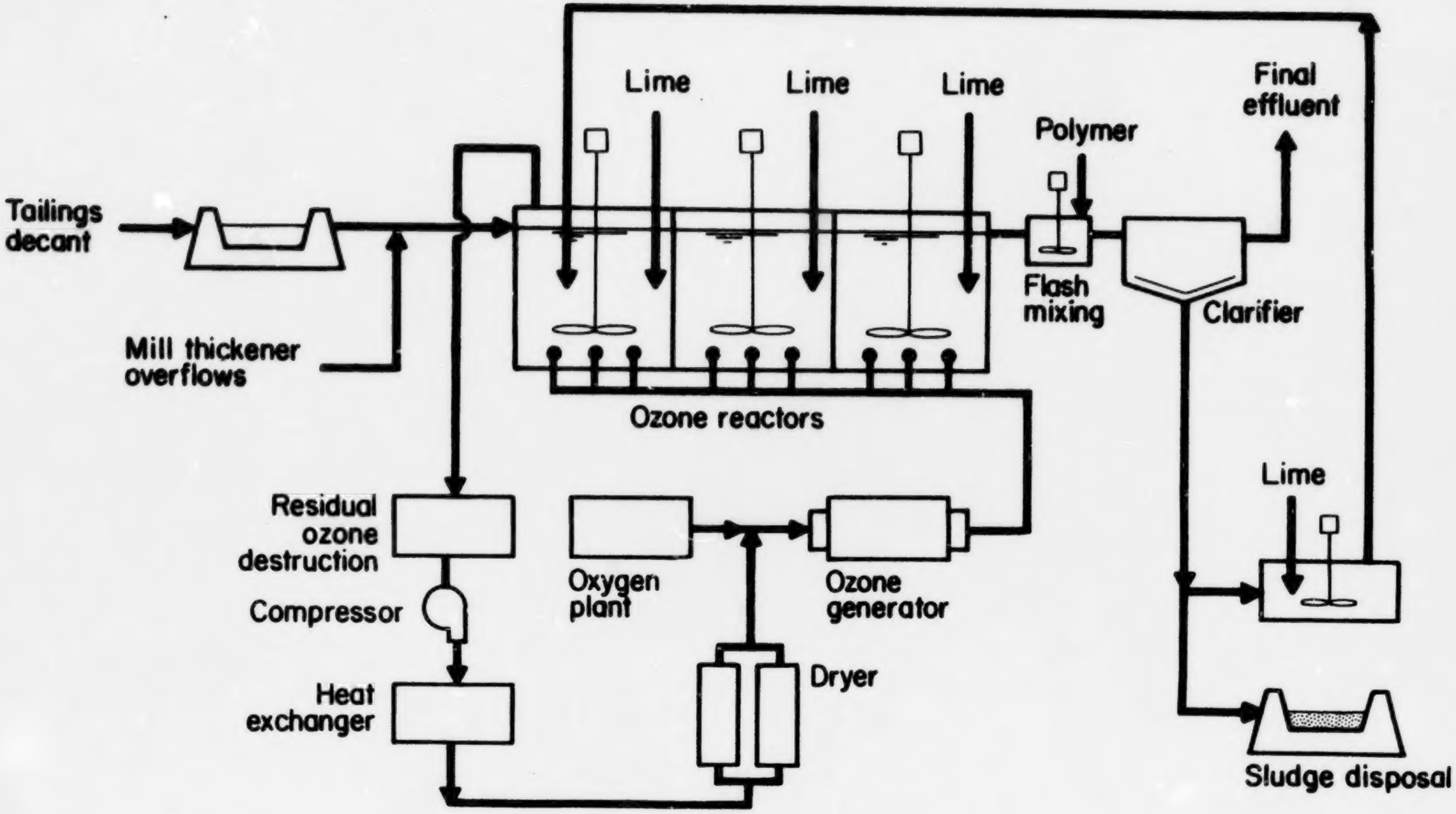

Figure 7 - Ozone Oxidation

The ozone system consists of an oxygen production plant, an ozone generator and residual ozone destruction facilities. Oxygen is supplied from a package liquifaction plant which feeds the gas directly to the ozone generators. There, ozone is produced as a 2% O_3 - 98% O_2 mixture which is injected into the reactors through diffusers. Additional mixing is provided by agitators to maintain efficient O_3/liquid contact. The oxygen remaining after the ozone reactions is recycled, with any residual ozone destroyed (to prevent corrosion problems in upstream equipment) by passing the oxygen through a thermal combustion unit. The O_2 is then compressed, cooled, dried, and returned to the ozone generators.

The principal process design parameters used for sizing the ozone oxidation facilities are summarized in Table XXVIII.

TABLE XXVIII

OZONE OXIDATION - PROCESS DESIGN PARAMETERS

Parameter	Design Basis		
	720 m^3/h. 24,500 kg/d S_2O_3	720 m^3/h. 12,300 kg/d S_2O_3	360 m^3/h. 12,300 kg/d S_2O_3
O_3 to S_2O_3 weight ratio requirements	1.1	1.1	1.1
Total reactor retention time (h)	1	1	1
No. of reactors	3	3	3
O_2 to O_3 weight ratio requirements	1.1	1.1	1.1
No of ozonators	30	15	15

Cost Estimates

The ozone oxidation system is highly capital intensive mainly due to the need for on-site oxygen and ozone generation facilities which account for roughly 80% of the total cost of the modules. Overall capital costs, estimated at $29 to $50.5 million for the three cases being considered, are among the highest of the treatment alternatives evaluated in this study.

Operating costs (mainly for power requirements for the ozone generation and destruction systems, and for oxygen generation and recycle) are related directly to the thiosalt loadings. As shown in Table XXIX, a reduction of 50% in thiosalt content reduces these costs from $8.5 million to $4.8 million/year.

The total annual costs (operating and annualized capital costs) are summarized in Table XXX.

TABLE XXX

OZONE OXIDATION COST SUMMARY

Item	Base Case[1.]	100% Flow 50% S_2O_3	50% Flow 50% S_2O_3
Annualized total cost (millions)	$18.54	$10.98	$10.58
Cost/m^3 water treated	$ 2.96	$ 1.73	$ 3.33
Cost/t ore milled	$ 5.08	$ 3.01	$ 2.89

1. 720 m^3/h, 24,500 kg/d thiosalts.

Comments of the Process

Ozone offers a number of advantages over other oxidants:

- It leaves no by-products (unlike chlorine) in the treated water;

TABLE XXIX - OZONE OXIDATION - COST ESTIMATES

	Base Case	100% Flow 50% Thio	50% Flow 50% Thio
Primary Module	$'000	$'000	$'000
Bare Equipment Items	594.00	594.00	440.00
Misc. Unspecified Items (15% of 1)	89.10	89.10	66.00
Subtotal Bare Equipment	683.10	683.10	506.00
Installed Cost (Bare Equipment x 1.45)	990.50	990.50	733.70
Other Items - Installed Cost	0.00	0.00	0.00
Subtotal Installed Equipment	990.50	990.50	733.70
Services: Electrical 15%, Instr. 10%, Serv. Piping 10%, Other Serv. 5%	396.20	396.20	293.48
Building 5,000.00 5,000.00 5,000.00	350.00	350.00	350.00
Total Primary Module Cost (A)	1,736.69	1,736.69	1,377.18
Ozone & Oxygen plant (B) (includes Installation & Services)	24,000.00	13,000.00	13,000.00
Settling Basin Cost (B)	150.00	150.00	100.00
Lime Module Cost (C)	2,990.00	2,650.00	2,160.00
Total Cost of Modules (A + B + C)	28,876.00	17,536.00	16,637.00
Engineering & Construction (25% of Modules)	7,219.00	4,384.00	4,159.00
Subtotal	36,095.00	21,920.00	20,796.00
Contingency (40% of Subtotal)	14,438.00	8,768.00	8,318.00
OVERALL TOTAL CAPITAL COST	50,533.00	30,688.00	29,114.00
Operating Cost	'000/y.	'000/y	'000/y
Materials, Supplies, Utilities & Labour	5,464.00	3,120.00	3,102.00
Lime	1,000.00	500.00	500.00
Maint., Taxes & Insurance (4.5% of Tot.Cap.)	1,974.79	1,223.51	1,152.68
TOTAL OPERATING COST	8,438.79	4,843.51	4,754.68

. As shown in Table XXVI, it is one of the strongest oxidants.

Because of the need to generate ozone directly for use, the treatment of industrial effluents with varying levels of contaminants requires stand-by capacity to provide sufficient ozone to meet peak requirements. The costs derived in the previous section are based on average thiosalt concentrations in the mill discharge. Actual thiosalt concentrations vary substantially, with maximum loadings being 50% higher than the mean. An ozone oxidation system would thus require stand-by capacity to meet these peak-loads. Alternately, an extensive equalizing system or treatment with other oxidants (e.g. H_2O_2, Cl_2) would be required. These options were not evaluated as a part of this study, but would substantially increase the capital costs.

The ozone equipment was sized by assuming O_3 to S_2O_3 weight requirements of 1.1 to 1. The CANMET laboratory studies have shown weight requirements varying over a wide range, and actual requirements may be higher than those assumed. This would result in both higher capital and operating costs.

The ozonation of milling effluents would occur quite readily. Discussions with ozone equipment manufacturers have indicated that at the large thiosalt concentrations expected, the reactions would be very rapid, and thus the reactors could be relatively simple in design.

Facilities required for the generation of oxygen and ozone are, however, quite complex. The City of Montreal ozonation system for the treatment of potable water (3500 kg/d O_3 capacity) has been plagued with numerous problems, with less than 20% available operation time since start-up in 1980. Thus, use of an ozone system may require a stand-by supply of other oxidants such as H_2O_2.

The largest ozone generating plants in the world are used currently for the treatment of potable water and produce 5000 kg/d O_3. For comparison, the ozone system required for the treatment of BMS effluents would produce 26,000 kg/d O_3.

3.2.3 Chlorine Oxidation

Chlorine is widely used for the treatment of potable water, sewage, and industrial effluents. The mining industry in the past decade has installed numerous chlorine oxidation plants for the treatment of both base metal and gold mill effluents containing cyanides. The reaction proposed for ths treatment of thiosalts by chlorine is as follows(1):

$$S_2O_3^{2-} + 4\,Cl_2 + 5H_2O \rightarrow 2SO_4^{2-} + 10H^+ + 8\,Cl^-$$

Based on this stoichiometry, for each kilogram of thiosalts oxidized, 2.54 kg Cl_2 and 2.64 kg $Ca(OH)_2$ are required. Bench-scale chlorine oxidation tests conducted by CANMET's Mineral Sciences Laboratories using synthetic thiosalt solutions have confirmed these requirements. More recent tests by both CANMET and the Noranda Research Centre using actual BMS thiosalt solutions have shown that chlorine reacts very rapidly with thiosalts. Reagent requirements were found however to be 70% higher than stoichiometric. Similar data have been observed during the treatment of gold milling effluents, where actual chlorine requirements are substantially higher than values predicted by stoichiometry.

A flowsheet of chlorine treatment for milling effluents is shown in Figure 8.

The decant from the tailings pond and mill thickener overflows are combined and sent for treatment in three stirred tank reactors operated in series at a total retention time of 45 min. Chlorine gas is sparged directly to the reactors with lime added to maintain the pH at ~9. Following chlorination, the effluent overflows to a flash mixing reactor where polymers are added to promote solid/liquid separation. Underflow sludge, made up mostly of gypsum and metal hydroxides, is sent for disposal in a sludge pond. The final effluent is discharged to the receiving water.

Oxidation of thiosalts by chlorine results in substantially higher lime requirements when compared to the use of other oxidants because of the initial formation of hypochlorite:

$$2\ Cl_2 + Ca\ (OH)_2 \rightarrow Ca\ (OCl)_2 + 2\ HCl$$

$$Ca\ (OH)_2 + 2\ HCl \rightarrow CaCl_2 + 2H_2O$$

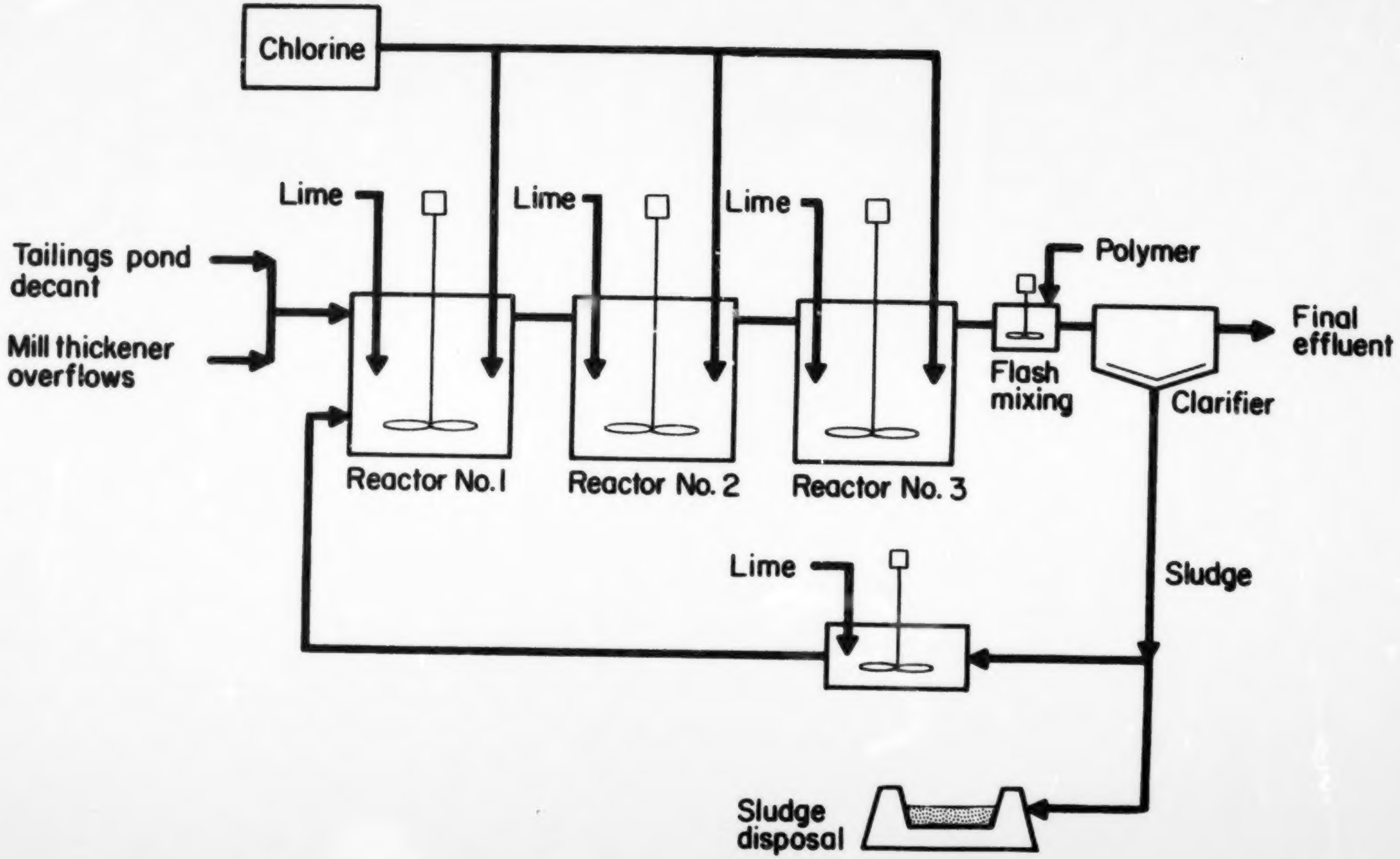

Figure 8 - Chlorine Oxidation

Because of this, an on-site kiln to generate lime from limestone was considered in addition to the usual lime module using quicklime.

The principal process design parameters used for sizing the chlorine oxidation treatment facilities are summarized in Table XXXI.

TABLE XXXI

CHLORINE OXIDATION - PROCESS DESIGN PARAMETERS

Parameter	Design Basis		
	720 m^3/h. 24,500 kg/d S_2O_3	720 m^3/h. 12,300 kg/d S_2O_3	360 m^3/h. 12,300 kg/d S_2O_3
Cl_2 to S_2O_3 weight ratio requirements	4.3	4.3	4.3
Total reactor retention time (h)	0.75	0.75	0.75
No. of reactors	3	3	3
Agitator power (kW/m^3)	0.17	0.17	0.17

Cost Estimates

Chlorination of thiosalt effluents requires a relatively simple flowsheet (Figure 8). Capital costs, as shown in Table XXXII are among the lowest of the systems evaluated, and range from $7.1 to $10.0 million for the three cases considered. An on-site kiln to generate lime would increase the capital costs significantly (e.g. for the base case to a total of $17.2 million (Table XXXIII)).

The operating costs are mainly due to chlorine ($7 to $13.8 million) and lime ($2.1 to $4.2 million). Although on-site

TABLE XXXII - CHLORINE OXIDATION - COST ESTIMATES

	Base Case	100% Flow 50% Thio	50% Flow 50% Thio
Primary Module	$'000	$'000	$'000
Bare Equipment Items	239.00	239.00	175.00
Misc. Unspectified Items (15% of 1)	35.85	35.85	26.25
Subtotal Bare Equipment	274.85	274.85	201.25
Installed Cost (Bare Equipment x 1.45)	398.53	398.53	291.81
Other Items - Installed Cost	0.00	0.00	0.00
Subtotal Installed Equipment	398.53	398.53	291.81
Services: Electrical 15%, Instr. 10%, Serv. Piping 10%, Other Serv. 5%	159.41	159.41	116.73b
Building 3,000.00 3,000.00 2,500.00	210.00	210.00	175.00
Total Primary Module Cost (A)	767.95	767.95	583.54
Settling Basin Cost (B)	150.00	150.00	100.00
Lime Module Cost (C)	4,800.00	3,770.00	3,350.00
Total Cost of Modules (A + B + C)	5,717.95	4,687.95	4,033.54
Engineering & Construction (25% of Modules)	1,429.49	1,171.99	1,008.38
Subtotal	7,147.43	5,859.93	5,041.92
Contingency (40% of Subtotal)	2,858.97	2,343.97	2,016.77
OVERALL TOTAL CAPITAL COST	10,006.40	8,203.90	7,058.69
Operating Cost	'000/y	'000/y	'000/y
Materials, Supplies, Utilities & Labour	13,769.00	7,037.00	7,003.00
Lime	4,160.00	2,080.00	2,080.00
Maint., Taxes & Insurance (4.5% of Tot.Cap.)	450.29	369.18	317.64
TOTAL OPERATING COST	18,379.29	9,486.18	9,400.64

TABLE XXXIII - CHLORINE OXIDATION (LIME KILN)
COST ESTIMATES

	Base Case
Primary Module	$'000
Bare Equipment Items	239.00
Misc. Unspecified Items (15% of 1)	35.85
Subtotal Bare Equipment	274.85
Installed Cost (Bare Equipment x 1.45)	398.53
Other Items - Installed Cost	0.00
Subtotal Installed Equipment	398.53
Services: Electrical 15%, Instr. 10%, Serv. Piping 10%, Other Serv. 5%	159.41
Building 3,000.00 3,000.00 2,500.00	210.00
Total Primary Module Cost (A)	767.95
Settling Basin Cost (B)	150.00
Lime Module Cost (C)	8,900.00
Total Cost of Modules (A + B + C)	9,817.95
Engineering & Construction (25% of Mudules)	2,454.49
Subtotal	12,272.43
Contingency (40% of Subtotal)	4,908.97
OVERALL TOTAL CAPITAL COST	17,181.40
Operating Cost	'000/y
Materials, Supplies, Utilities & Labour	13,769.00
Lime	1,664.00
Maint., Taxes & Insurance (4.5% of Tot.Cap.)	773.16
TOTAL OPERATING COST	16,206.16

lime production would reduce the reagent costs substantially, the saving would be nearly offset by an increase in capital costs. The total annual costs (operating and annualized capital costs) are summarized in Table XXXIV. As shown, annual costs range from \$10.8 to \$20.4 million. On-site lime generation lowers the annual costs by about \$800 k/y. The treatment charges correspond to costs of \$2.96 to \$5.59/t ore milled, and \$1.77 to \$3.43/m^3 water treated.

TABLE XXXIV

CHLORINE OXIDATION - COST SUMMARIES

Item	Base Case[1.]	100% Flow 50% S_2O_3	50% Flow 50% S_2O_3
Chlorination, Purchased Lime			
Annualized total cost (millions)	\$20.4	\$11.13	\$10.81
Cost/m^3 water treated	\$ 3.23	\$ 1.77	\$ 3.43
Cost/t ore milled	\$ 5.59	\$ 3.05	\$ 2.96
Chlorination, On-Site Lime Production			
Annualized total cost (millions)	\$19.64		
Cost/m^3 water treated	\$ 3.11		
Cost/t ore milled	\$ 5.39		

1. 720 m^3/h, 24,500 kg/d thiosalts.

Comments on the Process

Chlorination of milling effluents for the destruction of thiosalts is technically simple, with the basic flowsheet being similar to those currently used by the gold mining industry for the treatment of cyanide effluents. The equipment is familiar to mill operators (e.g. agitators, mix-tank reactors), and a full-scale facility based on currently available data could be installed with a high degree of certainty of successful operation.

The oxidation of thiosalts using chlorine results in the generation of 2.5g Cl^- for every g S_2O_3 destroyed. Thus, the reaction of 1400 mg/L S_2O_3 (base case) would result in the introduction of 3750 mg/L chloride. The possible detrimental effects of this discharge on biological activity in the receiving streams would have to be evaluated.

Unlike the use of ozone and H_2O_2, treatment of milling effluents using chlorine may result in the generation of chlorinated organic compounds due to the reactions of chlorine and organic mill reagents. Some chlorinated organic compounds are highly toxic, and additional studies would be required to evaluate potential detrimental effects.

3.2.4 SO_2-Air Oxidation

A process for the destruction of cyanide using catalyzed SO_2 - air oxidation has been developed by Inco Limited(^{19}). Possible reactions using SO_2-air for the destruction of thiosalts have been described previously, as follows(^{1}):

a) Formation of Caro's acid:

$$H_2O + SO_2 + O_2 \rightarrow H_2SO_5$$

b) Oxidation of thiosulphate by Caro's acid:

$$S_2O_3^{2-} + H_2O + 4H_2SO_5 + 2H^+ \rightarrow 6H_2SO_4$$

c) Neutralization of acid generated:

$$H_2SO_4 + Ca(OH)_2 \rightarrow CaSO_4 + 2H_2O$$

Laboratory tests to evaluate the SO_2-air process for the oxidation of thiosalts have been conducted by CANMET's Mineral Sciences Laboratories, Noranda Research Centre and BMS. CANMET found minimal changes in thiosalt levels in tests of up to 48-h duration but conducted without any copper catalyst addition. Noranda Research, using actual BMS solutions, and a range of conditions (pH 2 to 9, $SO_2:S_2O_3$ in excess, 100 mg/L Cu addition), also noted no significant reduction in thiosalt levels. Tests at BMS conducted in 1982 using laboratory flotation cells and fresh effluents produced complete thiosalt removal - the results were, however, uneven and inconsistent.

Based on the results of the BMS tests, a flowsheet for the treatment of milling effluents using the catalyzed SO_2- air oxidation process is shown in Figure 9.

The decant from the tailings pond and mill thickener overflows are combined, and sent for treatment in three stirred tank reactors operated in series at a total retention time of 60 min. A mix of SO_2 and air (at an approximate volumetric ratio of 1:20) is sparged into the tanks with the reactor agitators promoting efficient gas/liquid mass transfer. Copper sulphate (solution) is added to the first reactor. The pH is maintained at 8.5 to 9 using lime added to each of the reactors. The discharge from the third reactor overflows to a flash mixing chamber where flocculants are added. The liquid then passes to

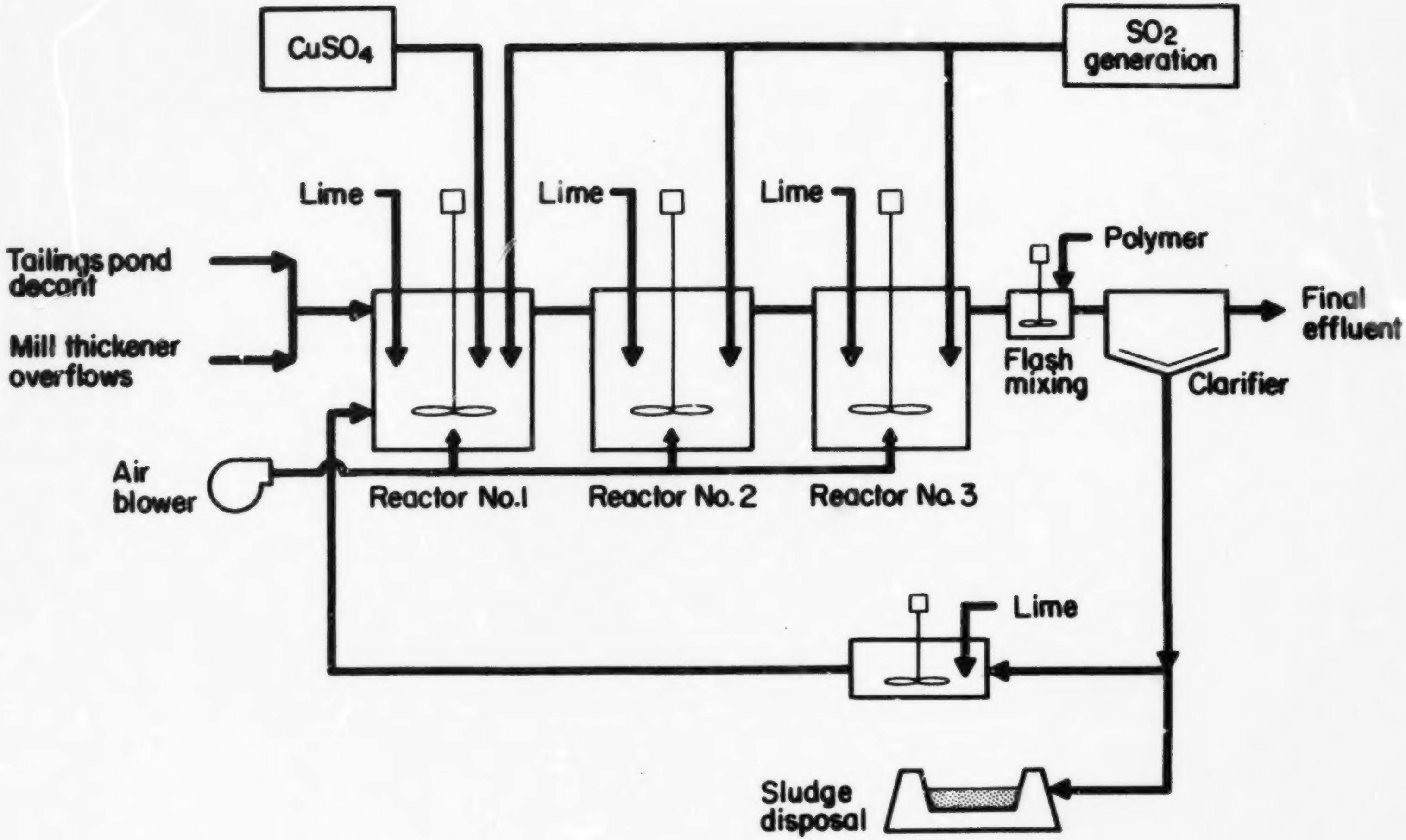

Figure 9 - SO_2 - Air Oxidation

a clarifier for solid/liquid separation. Underflow sludge is sent for disposal to the tailings pond, and the overflow is discharged to the receiving waters.

Because of the high SO_2 requirements, two methods for on-site generation of sulphur dioxide gas were considered; viz., burning of sulphur, and roasting of pyrite from the mill tailings. Lime requirements will also be high, as all the SO_2 gas added will be converted to H_2SO_4 which requires neutralization. Thus, the use of purchased quicklime and on-site lime production from limestone were considered.

The principal process parameters used for sizing the SO_2-air oxidation treatment facilities are summarized in Table XXXVI.

TABLE XXXV

SO_2 AIR OXIDATION - PROCESS DESIGN PARAMETERS

Parameter	Design Basis		
	720 m^3/h. 24,500 kg/d S_2O_3	720 m^3/h. 12,300 kg/d S_2O_3	360 m^3/h. 12,300 kg/d S_2O_3
SO_2 to S_2O_3 weight ratio requirements	2.3	2.3	2.3
Air addition, air/SO_2 volumetric ratio, %	2	2	2
Reactors - 2 lines of 3 each, total retention time (h)	1.5	1.5	1.5
Agitator Power (kW/m^3)	0.17	0.17	0.17

Cost Estimates

Estimates for the SO_2-air process were prepared assuming alternate sources of SO_2 gas and lime, as follows (Tables XXXVI to XXXIX):

- Sulphur burning, purchased quicklime: results in capital costs ranging from \$9.3 to \$13.3 million for the three cases considered. Annual operating costs vary from \$4.5 to \$8.6 million, with approximately 60% of the cost being due to lime.

- Sulphur burning, on-site lime production: results in greatly reduced lime costs (\$2.1 versus \$5.3 million for the base case). There is a corresponding increase in capital charges, however (from \$13.3 to \$22.0 million for the base case).

- Tailings roasting, purchased quicklime: The use of roasting facilities to produce SO_2 gas from pyrite contained in BMS tailings results in higher capital costs than the previous schemes, (overall costs range from \$11 million to \$15.5 million for the three cases considered with the use of quick lime).

- Tailings roasting, on-site lime production: While on-site lime production increases these costs still further, annual operating costs for SO_2 treatment with roasting of tailings are somewhat lower (\$4 to \$7.2 million), and on-site lime production lowers operating costs from \$7.2 to \$4.5 million/y for the base case.

TABLE XXXVI - SO_2 - AIR OXIDATION (SULPHUR)
COST ESTIMATES

	Base Case	100% Flow 50% Thio	50% Flow 50% Thio
Primary Module	$'000	$'000	$'000
Bare Equipment Items	880.00	687.00	560.00
Misc. Unspectified Items (15% of 1)	132.00	103.05	84.00
Subtotal Bare Equipment	1,012.00	790.05	644.00
Installed Cost (Bare Equipment x 1.45)	1,467.40	1,145.57	933.80
Other Items - Installed Cost	94.30	62.00	62.00
Subtotal Installed Equipment	1,561.70	1,207.57	995.80
Services: Electrical 15%, Instr. 10%, Serv. Piping 10%, Other Serv. 5%	624.68	483.03	398.32
Building 6,000.00 5,500.00 5,000.00	420.00	385.00	350.00
Total Primary Module Cost (A)	2,606.38	2,075.60	1,744.12
Settling Basin Cost (B)	150.00	150.00	100.00
Lime Module Cost (C)	4,850.00	3,850.00	3,450.00
Total Cost of Modules (A + B + C)	7,606.38	6,075.60	5,294.12
Engineering & Construction (25% of Modules)	1,901.60	1,518.90	1,323.53
Subtotal	9,507.98	7,594.50	6,617.65
Contingency (40% of Subtotal)	3,803.19	3,037.80	2,647.06
OVERALL TOTAL CAPITAL COST	13,311.17	10,632.30	9,264.71
Operating Cost	'000/y	'000/y	'000/y
Materials, Supplies, Utilities & Labour	2,735.00	1,521.00	1,475.00
Lime	5,300.00	2,650.00	2,650.00
Maint., Taxes & Insurance (4.5% of Tot.Cap.)	599.00	478.45	416.91
TOTAL OPERATING COST	8,634.00	4,649.45	4,541.91

TABLE XXXVII - AIR OXIDATION (SULPHUR/LIME KILN) - COST ESTIMATES

	Base Case
Primary Module	$'000
Bare Equipment Items	880.00
Misc. Unspectified Items (15% of 1)	132.00
Subtotal Bare Equipment	1,012.00
Installed Cost (Bare Equipment x 1.45)	1,467.40
Other Items - Installed Cost	94.30
Subtotal Installed Equipment	1,561.70
Services: Electrical 15%, Instr. 10%, Serv. Piping 10%, Other Serv. 5%	624.68
Building 6,000.00	420.00
Total Primary Module Cost (A)	2,606.38
Settling Basin Cost (B)	150.00
Lime Module Cost (C)	9,800.00
Total Cost of Modules (A + B + C)	12,556.38
Engineering & Construction (25% of Modules)	3,139.10
Subtotal	15,695.48
Contingency (40% of Subtotal)	6,278.19
OVERALL TOTAL CAPITAL COST	21,973.67
Operating Cost	'000/y
Materials, Supplies, Utilities & Labour	2,735.00
Lime	2,120.00
Maint., Taxes & Insurance (4.5% of Tot.Cap.)	988.81
TOTAL OPERATING COST	5,843.81

TABLE XXXVIII - SO$_2$ - AIR OXIDATION (PYRITE)
COST ESTIMATES

	Base Case	100% Flow 50% Thio	50% Flow 50% Thio
Primary Module	$'000	$'000	$'000
Bare Equipment Items	1,333.00	1,050.00	920.00
Misc. Unspecified Items (15% of 1)	199.95	157.50	138.00
Subtotal Bare Equipment	1,532.95	1,207.50	1,058.00
Installed Cost (Bare Equipment x 1.45)	2,222.78	1,750.88	1,534.10
Other Items - Installed Cost	141.50	93.00	93.00
Subtotal Installed Equipment	2,364.28	1,843.88	1,627.10
Services: Electrical 15%, Instr. 10%, Serv. Piping 10%, Other Serv. 5%	945.71	737.55	650.84
Building 8,000.00 7,500.00 7,000.00	560.00	525.00	490.00
Total Primary Module Cost (A)	3,869.99	3,106.43	2,767.94
Settling Basin Cost (B)	150.00	150.00	100.00
Lime Module Cost (C)	4,850.00	3,850.00	3,450.00
Total Cost of Modules (A + B + C)	8,869.99	7,106.43	6,317.94
Engineering & Construction (25% of Modules)	2,217.50	1,776.61	1,579.49
Subtotal	11,087.49	8.883.03	7,897.43
Contingency (40% of Subtotal)	4,434.99	3,553.21	3,158.97
OVERALL TOTAL CAPITAL COST	15,522.48	12,436.24	11,056.40
Operating Cost	'000/y	'000/y	'000/y
Materials, Supplies, Utilities & Labour	1,250.00	811.00	811.00
Lime	5,300.00	2,650.00	2,650.00
Maint., Taxes & Insurance (4.5% of Tot.Cap.)	698.51	559.63	497.54
TOTAL OPERATING COST	7,248.51	4,020.63	3,958.54

TABLE XXXIX - SO$_2$ AIR OXIDATION (PYRITE/LIME KILN) COST ESTIMATES

	Base Case
Primary Module	$'000
Bare Equipment Items	1,333.00
Misc. Unspecified Items (15% of 1)	199.95
Subtotal Bare Equipment	1,532.95
Installed Cost (Bare Equipment x 1.45)	2,222,78
Other Items - Installed Cost	141.50
Subtotal Installed Equipment	2,364.28
Services: Electrical 15%, Instr. 10%, Serv. Piping 10%, Other Serv. 5%	945.71
Building 8,000.00	560.00
Total Primary Module Cost (A)	3,869.99
Settling Basin Cost (B)	150.00
Lime Module Cost (C)	9,800.00
Total Cost of Modules (A + B + C)	13,819.99
Engineering & Construction (25% of Modules)	3,455.00
Subtotal	17,274.99
Contingency (40% of Subtotal)	6,909.99
OVERALL TOTAL CAPITAL COST	24,184.98
Operating Cost	'000/y
Materials, Supplies, Utilities & Labour	1,250.00
Lime	2,120.00
Maint., Taxes & Insurance (4.5% of Tot.Cap.)	1,088.32
TOTAL OPERATING COST	4,458.32

The total annual costs (operating and annualized capital costs) are summarized in Table XL. On-site lime production and SO_2 generation by roasting of tailings result in the lowest costs ($9.3 million/y for base case, versus $10.2 to $11.3 million for the other systems). This translates into $2.55/t ore milled, or $1.47/$m^3$ water treated.

Comments on the Process

Unlike the other chemical oxidants considered (i.e. H_2O_2, O_3, Cl_2) the use of catalyzed SO_2-air oxidation is technically uncertain, and extensive test work would be required to confirm the effectiveness of the process for the destruction of thiosalts. The results of laboratory tests conducted to date have shown inconsistent results although the reactions with SO_2-air appear to be significantly slower than the corresponding reactions with chlorine or O_3.

The use of SO_2 results in substantially higher generation of acidity as compared to H_2O_2 or ozone. In addition to the production of H_2SO_4 from thiosalts, all the SO_2 used in the process is converted to H_2SO_4, and subsequently requires neutralization.

The oxidation of thiosalts with SO_2-air would not result in other contaminants being introduced into the final effluent. The SO_2 added would be converted to H_2SO_4, and would be precipitated by lime as gypsum. Copper which is added to catalyze the reaction, would also be precipitated with the gypsum as copper oxide at the operational pH of 8.5 to 9.

TABLE XL

SO_2 - AIR OXIDATION COST SUMMARIES

Item	Base Case[1.]	100% Flow 50% S_2O_3	50% Flow 50% S_2O_3
Sulphur/Purchased Lime			
Annualized total cost (millions)	$11.30	$ 6.78	$ 6.39
Cost/m^3 water treated	$ 1.79	$ 1.07	$ 2.02
Cost/t ore milled	$ 3.10	$ 1.86	$ 1.75
Sulphur/On-Site Lime			
Annualized total cost (millions)	$10.2		
Cost/m^3 water treated	$ 1.63		
Cost/t ore milled	$ 2.81		
Pyrite/Purchased Lime			
Annualized total cost (millions)	$10.35	$ 6.51	$ 6.17
Cost/m^3 water treated	$ 1.65	$ 1.03	$ 1.95
Cost/t ore milled	$ 2.83	$ 1.78	$ 1.68
Pyrite/On-Site Lime			
Annualized total cost (millions)	$ 9.30		
Cost/m^3 water treated	$ 1.47		
Cost/t ore milled	$ 2.55		

1. 720 m^3/h, 24,500 kg/d thiosalts

3.3 Electrochemical Oxidation of Thiosalts

The electrochemical oxidation of thiosalt solutions was studied in laboratory scale tests at the Noranda Research Centre(^{20}) using a variety of electrolytic cells. Results of the tests indicated that 80% of the thiosalts could be fully oxidized to SO_4. The remaining 20% were converted to dithionate, S_2O_6, which could not be oxidized anodically. Power requirements are influenced by solution concentrations, with better efficiencies (i.e. lower power and reduced electrode surface area requirements) being observed at higher thiosalt concentrations. A flowsheet of a potential electrochemical facility for the treatment of thiosalts in milling effluents is presented in Figure 10.

As indicated, mill tailings decant and thickener overflows are passed through a settling pond for settleable solids removal. Subsequently, a sand filter is used to remove most of the remaining solids in order to minimize potential solids interference during electrochemical treatment. The electrochemical reactor consists of standard modular electrolytic cells. For the treatment of 24,000 kg/d thiosalts, it has been estimated that 20 groups of cells with 6 cells per group, would be needed. Total electrode surface area required would be 9000 m^2. The process design parameters used in sizing the electrochemical facilities are summarized in Table XLI.

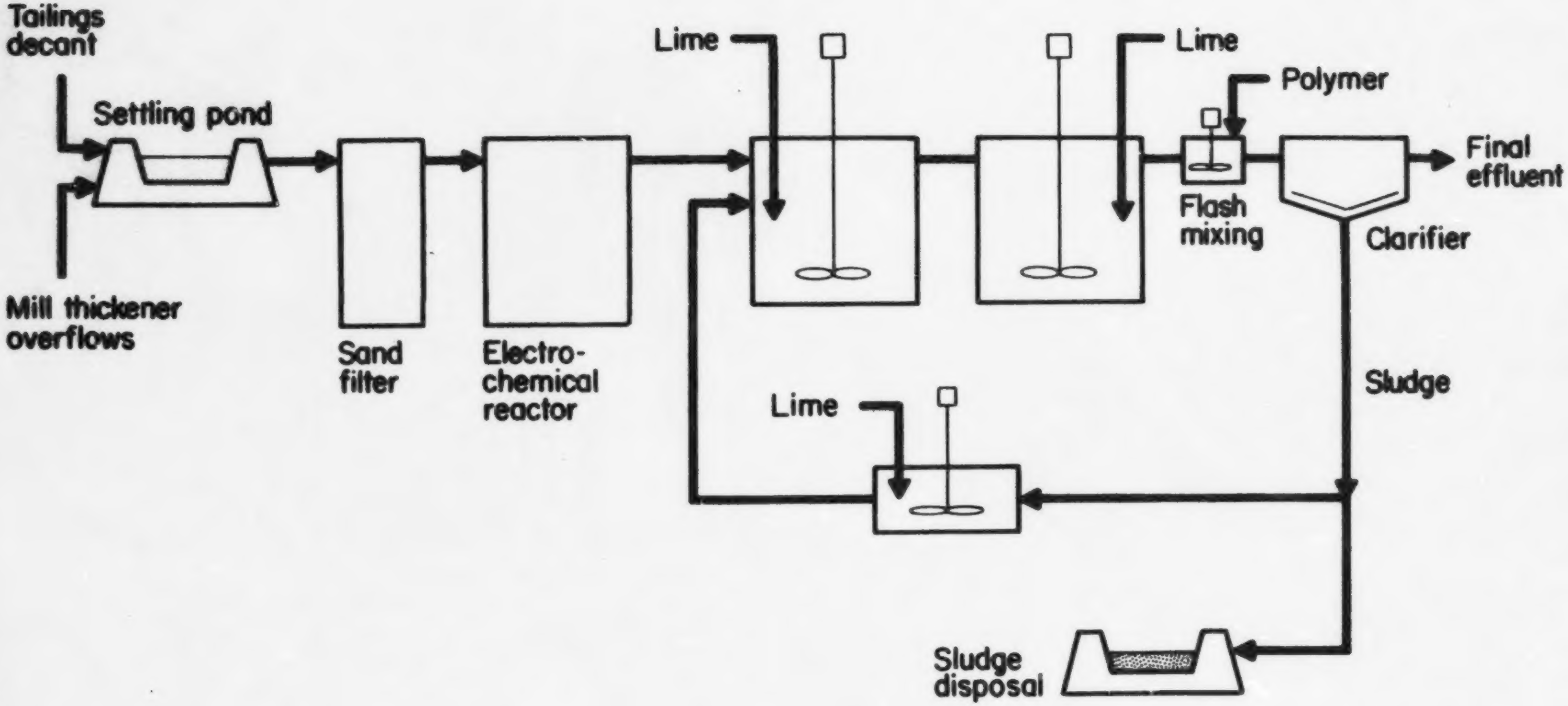

Figure 10 - Electrochemical Oxidation

TABLE XLI
ELECTROCHEMICAL OXIDATION - PROCESS DESIGN PARAMETERS

	Design Basis		
Parameter	12,000 L/min. 24,000 kg/d S_2O_3	12,000 L/min. 12,000 kg/d S_2O_3	6,000 L/min. 12,000 kg/d S_2O_3
1. Current density (MA/m^2)	20	15	20
2. Electrode area	9000	6000	4500
3. Number of cells	120	90	60
4. Power requirements, kWh/kg S_2O_3	15	20	15
5. Sandfilter (m^2)	75	75	40

Cost Estimates

As shown in Table XLII, overall capital costs for electrochemical treatment range from $45.3 to $79 million. The major portion of the cost is accounted for by the electrolytic equipment. The operating charges, mainly for electric power, range from $8.9 to $17 million per year. The total annual costs (operating and annualized capital costs) summarized in Table XLIII vary from $17.9 million to $32.9 million. This corresponds to $4.91 to $8.99/t ore milled and $3.49 to $5.68/$m^3$ water treated.

TABLE XLIII
ELECTROCHEMICAL OXIDATION COST SUMMARIES

Item	Base Case[1.]	100% Flow 50% S_2O_3	50% Flow 50% S_2O_3
Annualized total cost (millions)	$32.85	$22.07	$17.94
Cost/m^3 water treated	$ 5.20	$ 3.49	$ 5.68
Cost/t ore milled	$ 8.99	$ 6.03	$ 4.91

1. 720 m^3/h, 24,500 kg/d thiosalts.

TABLE XLII - ELECTROCHEMICAL OXIDATION
COST ESTIMATE

	Base Case	100% Flow 50% Thio	50% Flow 50% Thio
Primary Module	$'000	$'000	$'000
Bare Equipment Items	0.00	0.00	0.00
Misc. Unspectified Items (15% of 1)	0.00	0.00	0.00
Subtotal Bare Equipment	0.00	0.00	0.00
Installed Cost (Bare Equipment x 1.45)	0.00	0.00	0.00
Other Items - Installed Cost	1,000.00	1,000.00	660.00
Subtotal Installed Equipment	1,000.00	1,000.00	660.00
Services: Electrical 15%, Instr. 10%, Serv. Piping 10%, Other Serv. 5%	400.00	400.00	264.00
Building 0.00 0.00 0.00	0.00	0.00	0.00
Total Primary Module Cost (A)	1,400.00	1,400.00	924.00
Electrochemical Module (B) (includes bldg., installation & services)	40,000.00	26,700.00	22,000.00
Settling Basin Cost (B)	150.00	150.00	100.00
Lime Module Cost (C)	3,580.00	3,240.00	2,850.00
Total Cost of Modules (A + B + C)	45,130.00	31,490.00	25,874.00
Engineering & Construction (25% of Modules)	11,282.50	7,872.50	6,468.50
Subtotal	56,412.50	39,362.50	32,342.50
Contingency (40% of Subtotal)	22,565.00	15,745.00	12,937.00
OVERALL TOTAL CAPITAL COST	78,977.50	55,107.50	45,279.50
Operating Cost	'000/y	'000/y	'000/y
Materials, Supplies, Utilities & Labour	12,500.00	8,067.00	6,350.00
Lime	1,000.00	500.00	500.00
Maint., Taxes & Insurance (4.5% of Tot.Cap.)	3,553.99	2,479.84	2,037.58
TOTAL OPERATING COST	17,053.99	11,046.84	8,887.58

Comments on the Process

There are a number of technical uncertainties regarding the application of electrochemical treatment for milling effluents. As indicated in Tables IV and V, these effluents contain high concentrations of dissolved contaminants which may lead to the formation of precipitates on the electrodes. Also, complex multipath chemical and electrochemical reactions occur during electrolytic oxidation, and these can result in the decomposition of part of the thiosalts to elemental sulphur, and in potential contamination of the electrodes.

Thiosalt concentrations in milling effluents vary widely, from <200 mg/L to >1500 mg/L S_2O_3. The electrochemical treatment of low level thiosalt streams (<500 mg/L) would result in a great reduction in current efficiency, thereby requiring significant increases in power and reactor retention times which could increase costs significantly.

Dithionates are formed during anodic oxidation of thiosalts, with production ranging from 8 to 20% of the final total sulphate concentration. The effects of the release of dithionates to the receiving waters (e.g. potential of oxidation to SO_4) have not been studied, but certainly cannot be ignored.

Finally, at the present time, electrochemical treatment of effluents is generally limited to specific streams with low flows and high contaminant concentrations. The thiosalt effluent flows are very high, and as the costs indicate, these high flows result in very high capital and operating charges.

3.4 Reverse Osmosis

The reverse osmosis (RO) process involves forcing water under pressure (commonly 2000 to 10,000 kPa) through a semi-permeable membrane such as cellulose acetate or polyamide. This results in two streams; a purified product water, and a concentrated solution.

The process has been previously evaluated for the treatment of acid mine drainage(21), and is used widely for brackish water purification(22). When pilot tested on acid mine drainage, RO provided effective treatment without major iron or calcium sulphate scaling problems at 75% purified water recovery. At 85% recovery, however, severe scaling due to calcium sulphate precipitation took place. Also, oxidation of ferrous ions by bacteria occurred in the membranes, requiring pre-treatment by ultra-violet disinfection and/or lowering the pH to 2.9.

Preliminary bench-scale tests using both synthetic and actual BMS-M mill effluents were conducted by CANMET's Mineral Sciences Laboratories(1), and these experiments yielded product water with thiosalt concentrations below 100 mg/L (feed levels ranged from 900 to 4000 mg/L S_2O_3).

A flowsheet for the treatment of milling effluents by the reverse osmosis process is illustrated in Figure 11. Since particulates can cause membrane fouling, mill effluent streams (tailings decant, thickener overflows) pass initially through two-step solid/liquid separation consisting of a settling pond and a sand-filter. To prevent scale formation on the membrane

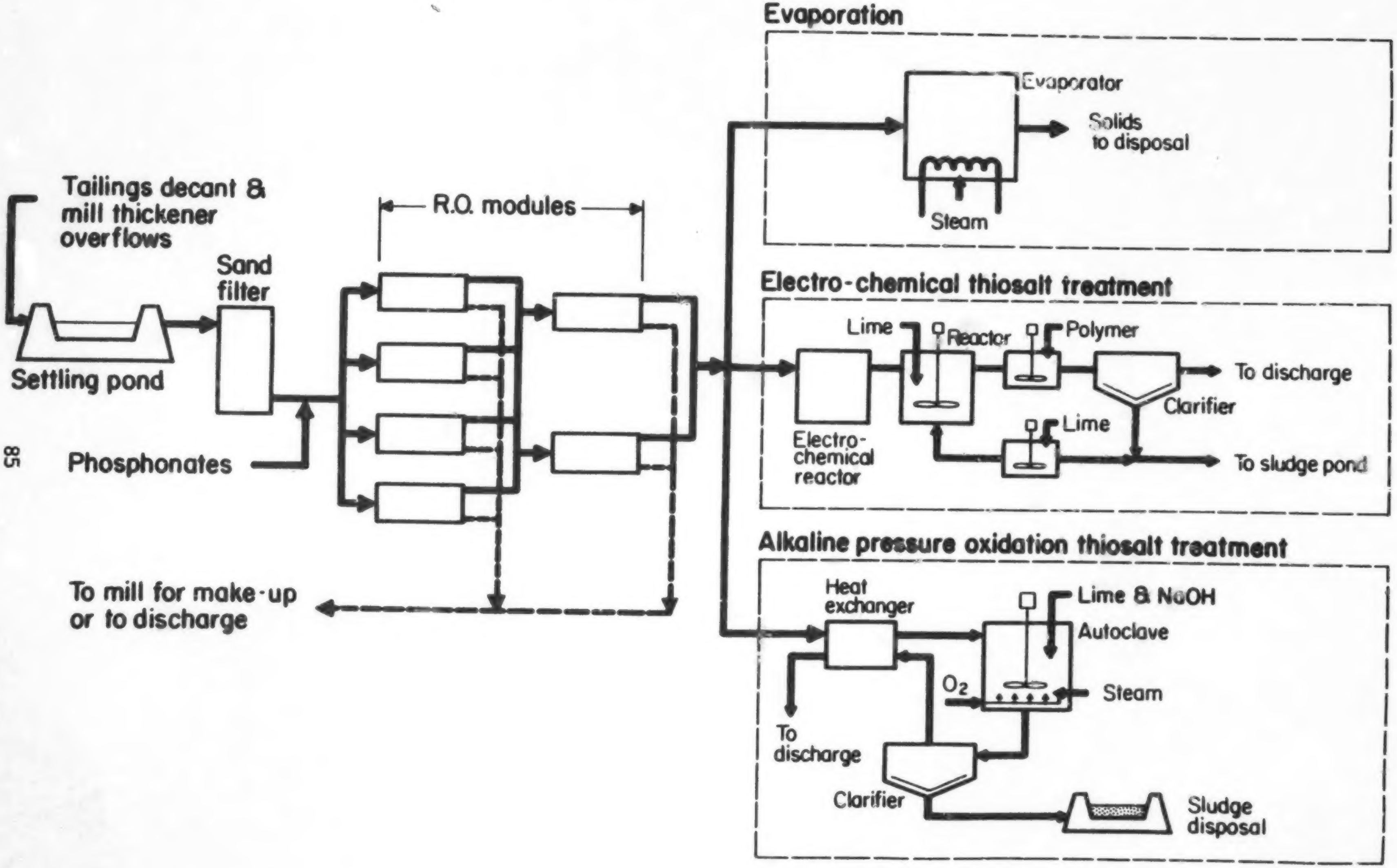

Figure 11 - Treatment by Reverse Osmosis

material, anti-scaling reagents are injected into the feed stream prior to RO treatment. The reverse osmosis system consists of two stages of skid-mounted units. Product water, expected to contain less than 100 mg/L of thiosalts can be discharged to the receiving waters or recycled to the mill for process use. The concentrated solution, containing >90% of the thiosalts, requires additional treatment.

Three treatment alternatives for the concentrated solution were evaluated:

- Evaporation, with disposal of the contaminant solids to a secured landfill.
- Electrochemical treatment using the flowsheet considered in section 3.3.
- Alkaline pressure oxidation using a flowsheet described in section 3.5.

The principal process design parameters used for sizing the reverse osmosis facilities are summarized in Table XLIV.

TABLE XLIV
REVERSE OSMOSIS - PROCESS DESIGN PARAMETERS

Parameter	Design Basis		
	12,000 L/min. 24,000 kg/d S_2O_3	12,000 L/min. 12,000 kg/d S_2O_3	6,000 L/min. 12,000 kg/d S_2O_3
1. Sand filter area (m^2)	75	75	40
2. Reverse osmosis system recovery (%)	75	75	75
3. Product water recovery (%)	75	75	75
4. Brine/feed ratio (%)	25	25	25
5. Evaporator sizing ($L/min/m^2$)	2.0	2.0	2.0
6. Evaporator surface area (m^2)	1500	1500	750
7. Volume evaporated (L/min)	3000	3000	1500

Cost Estimates

Estimates for the treatment of milling effluents by reverse osmosis, followed by three methods of concentrated solution disposal, are summarized in Tables XLV to XLVII. RO/concentrated solution evaporation is lowest in capital cost, with a range of $14.5 to $22.2 million for the three cases considered. Both alkaline pressure oxidation and electrochemical oxidation are substantially costlier, with capital requirements of $22.5 to $33.5 million and $26.9 to $44.7 million, respectively.

The cost to operate an RO/concentrated solution evaporation system was lower at the base (720 m^3/h, 24,500 kg/d S_2O_3) and 50% flow, 50% S_2O_3 conditions, with annual expenditures of $10.6 and $5.7 million, respectively. At 100% flow, 50% S_2O_3 conditions, RO/electrochemical treatment proved less costly.

The total annual expenditures (operating and amortized capital) are summarized in Table XLVIII. Lowest costs generally resulted for RO/evaporation systems, with total annual costs ranging from $8.6 to $15.1 million. This corresponds to costs of $2.37 to $4.13/t ore milled, and $2.38 to $2.71/$m^3$ water treated.

Comments on the Process

As indicated above, the reverse osmosis process does not result in the destruction of thiosalts. Rather, the thiosalts are concentrated (roughly four-fold), and the solution then has to be treated. There are a number of factors to

TABLE XLV - REVERSE OSMOSIS/EVAPORATION
COST ESTIMATES

	Base Case	100% Flow 50% Thio	50% Flow 50% Thio
Primary Module	$'000	$'000	$'000
Bare Equipment Items	930.00	930.00	625.00
Misc. Unspectified Items (15% of 1)	139.50	139.50	93.75
Subtotal Bare Equipment	1,069.50	1,069.50	718.75
Installed Cost (Bare Equipment x 1.45)	1,550.78	1,550.78	1,042.19
Other Items - Installed Cost	7,200.00	7,200.00	4,660.00
Subtotal Installed Equipment	8,750.78	8,750.78	5,702.19
Services: Electrical 15%, Instr. 10%, Serv. Piping 10%, Other Serv. 5%	3,500.31	3,500.31	2,280.88
Building 4,000.00 4,000.00 3,000.00	280.00	280.00	210.00
Total Primary Module Cost (A)	12,531.09	12,531.09	8,193.06
Settling Basin Cost (B)	150.00	150.00	100.00
Lime Module Cost (C)	0.00	0.00	0.00
Total Cost of Modules (A + B + C)	12,681.09	12,681.09	8,293.06
Engineering & Construction (25% of A+B+C)	3,170.27	3,170.27	2,073.27
Subtotal	15,851.36	15,851.36	10,366.33
Contingency (40% of Subtotal)	6,340.54	6,340.54	4,146.53
OVERALL TOTAL CAPITAL COST	22,191.90	22,191.90	14,512.86
Operating Cost	'000/y	'000/y	'000/y
Materials, Supplies, Utilities & Labour	9,633.00	9,633.00	5,015.00
Lime	0.00	0.00	0.00
Maint., Taxes & Insurance (4.5% of Tot.Cap.)	998.64	998.64	653.08
TOTAL OPERATING COST	10,631.64	10,631.64	5,668.08

TABLE XLVI - REVERSE OSMOSIS/ALKALINE PRESSURE OXIDATION - COST ESTIMATES

	Base Case	100% Flow 50% Thio	50% Flow 50% Thio
Primary Module	$'000	$'000	$'000
Bare Equipment Items	30.00	30.00	30.00
Misc. Unspectified Items (15% of 1)	4.50	4.50	4.50
Subtotal Bare Equipment	34.50	34.50	34.50
Installed Cost (Bare Equipment x 1.45)	50.03	50.03	50.03
Other Items - Installed Cost	7,200.00	7,200.00	4,660.00
Subtotal Installed Equipment	7,250.03	7,250.03	4,710.03
Services: Electrical 15%, Instr. 10%, Serv. Piping 10%, Other Serv. 5%	2,900.01	2,900.01	1,884.01
Building 4,000.00 4,000.00 3,000.00	280.00	280.00	210.00
Total Primary Module Cost (A)	10,430.04	10,430.04	6,804.04
Alk.Oxid. Moudle (incl. install. & services) (B)	3,000.00	3,000.00	2,100.00
Oxygen Plant (includes bldg. & Services) (C)	3,600.00	2,400.00	2,400.00
Settling Basin Cost (D)	150.00	150.00	100.00
Lime Module Cost (E)	1,980.00	1,650.00	1,440.00
Total Cost of Modules (A+B+C+D+E)	19,160.04	17,630.04	12,844.04
Engineering and Construction (25% of Modules)	4,790.01	4,407.51	3,211.01
Subtotal	23,950.04	22,037.54	16,055.04
Contingency (40% of Subtotal)	9,580.02	8,815.02	6,422.02
Washed Rock (Includes 20% Contingency)			
OVERALL TOTAL CAPITAL COST	33,530.06	30,852.56	22,477.06
Operating Cost	'000/y	'000/y	'000/y
Materials, Supplies, Utilities & Labour	9,175.00	8,351.00	4,788.00
Lime	1,000.00	500.00	500.00
Maint., Taxes & Insurance (4.5% of Tot.Cap.)	1,508.85	1,388.37	1,011.47
TOTAL OPERATING COST	11,683.85	10,239.37	6,299.47

TABLE XLVII - REVERSE OSMOSOS/ELECTROCHEMICAL
OXIDATION - COST ESTIMATES

	Base Case	100% Flow 50% Thio	50% Flow 50% Thio
Primary Module	$'000	$'000	$'000
Bare Equipment Items	30.00	30.00	30.00
Misc. Unspectified Items (15% of 1)	4.50	4.50	4.50
Subtotal Bare Equipment	34.50	34.50	34.50
Installed Cost (Bare Equipment x 1.45)	50.03	50.03	50.03
Other Items - Installed Cost	7,200.00	7,200.00	4,660.00
Subtotal Installed Equipment	7,250.03	7,250.03	4,710.03
Services: Electrical 15%, Instr. 10%, Serv. Piping 10%, Other Serv. 5%	2,900.01	2,900.01	1,884.01
Building 3,000.00 3,000.00 2,500.00	210.00	210.00	175.00
Total Primary Module Cost (A)	10,360.04	10,360.04	6,769.04
Elec.Chem. Module (incl.bldg. inst.& services)(B)	13,000.00	8,000.00	7,000.00
Oxygen Plant (includes bldg. & services)(C)	0.00	0.00	0.00
Settling Basin Cost (D)	150.00	150.00	100.00
Lime Module Cost (E)	2,020.00	1,700.00	1,490.00
Total Cost of Modules (A+B+C+D+E)	25,530.04	20,210.04	15,359.04
Engineering & Construction (25% of Modules)	6,382.51	5,052.51	3,839.76
Subtotal	31,912.54	25,262.54	19,198.79
Contingency (40% of Subtotal)	12,765.02	10,105.02	7,679.52
OVERALL TOTAL CAPITAL COST	44,677.56	35,367.56	26,878.31
Operating Cost	'000/y	'000/y	'000/y
Materials, Supplies, Utilities & Labour	9,353.00	7,353.00	4,877.00
Lime	1,000.00	500.00	500.00
Maint., Taxes & Insurance (4.5% of Tot.Cap.)	2,010.49	1,591.54	1,209.52
TOTAL OPERATING COST	12,363.49	9,444.54	6,586.52

TABLE XLVIII

REVERSE OSMOSIS COST SUMMARIES

Item	Base Case[1]	100% Flow 50% S_2O_3	50% Flow 50% S_2O_3
RO/Evaporation			
Annualized total cost (millions)	$15.07	$15.07	$ 8.57
Cost/m^3 water treated	$ 2.38	$ 2.39	$ 2.71
Cost/t ore milled	$ 4.13	$ 4.13	$ 2.37
RO/Electrochemical			
Annualized total cost	$21.30	$16.52	$11.96
Cost/m^3 water treated	$ 3.43	$ 2.62	$ 3.78
Cost/t ore milled	$ 5.83	$ 4.53	$ 3.27
RO-Alkaline Pressure Oxidation			
Annualized total cost	$18.39	$16.41	$10.79
Cost/m^3 water treated	$ 2.91	$ 2.59	$ 3.41
Cost/t ore milled	$ 5.03	$ 4.50	$ 2.95

1. 720 m^3/h, 24,500 kg/d thiosalts.

consider regarding the potential application of reverse osmosis for the treatment of milling effluents:

a) Even with the use of descaling agents, membrane life may be limited because of the high probability of scaling.

b) Thiosalt-oxidizing bacteria may build-up in the reverse-osmosis system (expecially tje membranes), and this could lead to pH depression and equipment corrosion problems.

c) RO followed by evaporation appears most cost-effective. Evaporation is a highly energy intensive process, with future operating costs dependent on the energy situation.

d) The facilities are more complex than chemical oxidation systems, and will require better training of operators to maintain satisfactory performance.

e) Extensive test-work would be required to define design parameters for a full-scale facility using RO.

3.5 Alkaline Pressure Oxidation

CANMET's Mineral Sciences Laboratories have developed a process for the oxidation of thiosalts in mill effluent using alkaline pressure oxidation(1). In laboratory scale experiments using synthetic thiosalt solutions, complete thiosalt oxidation was achieved using oxygen at retention times of less than 15 min. at a temperature of ~100°C and a pressure of 2000 kPa. A flowsheet for the oxidation of thiosalts in milling effluents using this process is presented in Figure 12.

The mill tailings decant is pumped to one of two settling ponds for settleable solids removal. Then this stream

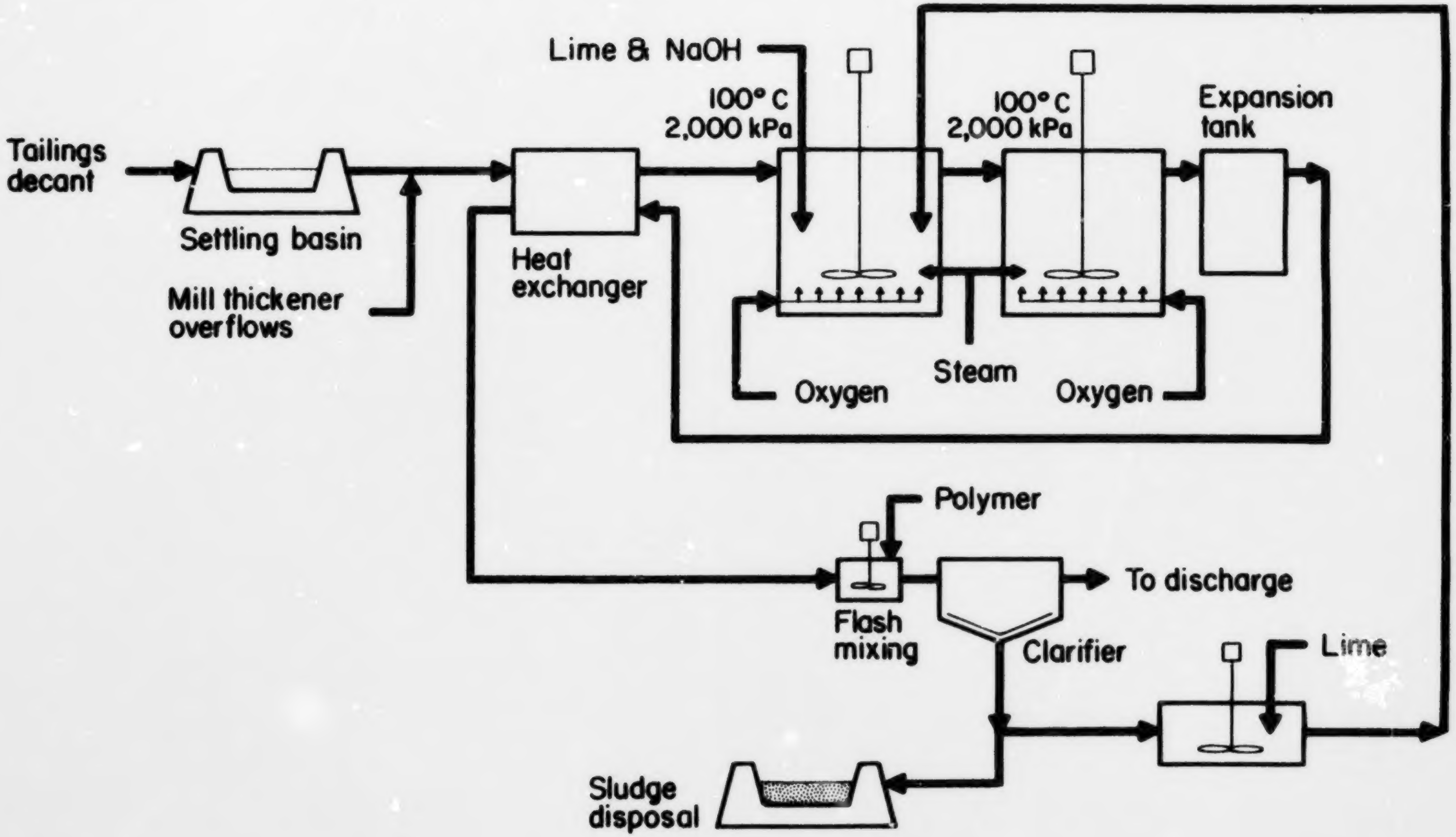

Figure 12 - Alkaline Pressure Oxidation

and warm thickener overflows from the mill are sent to a heat exchanger for pre-heating using waste heat recovered from the treated effluent. Subsequently, the combined stream is sent to alkaline pressure oxidation in two autoclaves, operated in series, each with a retention time of 15 minutes. In bench-scale tests, the use of lime alone for pH control resulted in incomplete oxidation; thus, lime and sodium hydroxide are added in the ratio of 2.2:1 $Ca(OH)_2$ to NaOH to neutralize SO_4 formed. The autoclaves are maintained at 2000 kPa with pure oxygen being added through spargers in the autoclaves. Agitators are used to promote efficient oxygen transfer, heat transfer and mixing in the reactors. The temperature in the autoclaves is maintained at 100°C with steam.

Following thiosalt oxidation, the effluent is brought to atmospheric pressure in an expansion tank, and then is passed through heat exchangers to pre-heat incoming effluent. From the heat exchangers, the stream is sent through a flash mixing tank, with flocculants added to promote settling, and then to a clarifier for solid/liquid separation. Sludges (mostly gypsum) are sent to a sludge disposal pond, and the clean overflow is discharged to the receiving waters.

The principal process design parameters used for sizing the alkaline pressure oxidation facilities are summarized in Table XLIX.

TABLE XLIX

ALKALINE PRESSURE OXIDATION - PROCESS DESIGN PARAMETERS

Parameter	Design Basis		
	720 m^3/h. 24,500 kg/d S_2O_3	720 m^3/h. 12,300 kg/d S_2O_3	360 m^3/h. 12,300 kg/d S_2O_3
O_2 to S_2O_3 weight ratio requirements	0.86	0.86	0.86
Alkaline requirements:			
NaOH to S_2O_3 weight ratio	0.4	0.4	0.4
$Ca(OH)_2$ to S_2O_3 weight ratio	0.8	0.8	0.8
Operating temp ($^\circ$C)	100	100	100
Autoclave total retention time (h)	0.5	0.5	0.5
Autoclave volume (m^3)	180	180	90
No. of autoclaves	2	2	2

Cost Estimates

Capital costs for the alkaline pressure oxidation process are highly flow dependent, ranging from \$16.3 to \$23.7 million for the three cases under consideration (Table L). Major capital items, as shown in Table B-13, Appendix B, include on-site oxygen generation facilities, autoclaves, and a steam plant. Operating costs, dominated by steam requirements, are proportional to the effluent flows treated, with costs ranging from \$6.5 million to \$12.5 million. The total annual costs (operating and annualized capital) are summarized in Table LI, and amount to \$9.7 to \$17.3 million. This corresponds to \$2.68 to \$4.73/t ore milled, and \$2.44 to \$3.08/m^3 water treated.

TABLE L - ALKALINE PRESSURE OXIDATION
COST ESTIMATES

	Base Case	100% Flow 50% Thio	50% Flow 50% Thio
Primary Module	$'000	$'000	$'000
Bare Equipment Items	25.00	25.00	25.00
Misc. Unspectified Items (15% of 1)	3.75	3.75	3.75
Subtotal Bare Equipment	28.75	28.75	28.75
Installed Cost (Bare Equipment x 1.45)	41.69	41.69	41.69
Other Items - Installed Cost	0.00	0.00	0.00
Subtotal Installed Equipment	41.69	41.69	41.69
Services: Electrical 15%, Instr. 10%, Serv. Piping 10%, Other Serv. 5%	16.68	16.68	16.68
Building 2,500.00 2,500.00 2,000.00	175.00	175.00	140.00
Total Primary Module Cost (A)	233.36	233.36	198.36
Alk. Oxid.Module (incl. install.& services)(B)	6,600.00	6,600.00	4,400.00
Oxygen Plant (includes bldg & services) (C)	3,600.00	2,400.00	2,400.00
Settling Basin Cost (D)	150.00	150.00	100.00
Lime Module Cost (E)	2,950.00	2,620.00	2,230.00
Total Cost of Modules (A+B+C+D+E)	13,533.36	12,003.36	9,328.36
Engineering & Construction (25% of Modules)	3,383.34	3,000.84	2,323.09
Subtotal	16,916.70	15,004.20	11,660.45
Contingency (40% of Subtotal)	6,766.68	6,001.68	4,664.18
OVERALL TOTAL CAPITAL COST	23,683.38	21,005.88	16,324.63
Operating Cost	'000/y	'000/y	'000/y
Materials, Supplies, Utilities & Labour	10,667.00	9,844.00	5,335.00
Lime	800.00	400.00	400.00
Maint., Taxes & Insurance (4.5% of Tot.Cap.)	1,065.75	945.26	734.61
TOTAL OPERATING COST	12,532.75	11,189.26	6,469.61

TABLE LI

ALKALINE PRESSURE OXIDATION COST SUMMARIES

Item	Base Case[1.]	100% Flow 50% S_2O_3	50% Flow 50% S_2O_3
Annualized total cost (millions)	$17.27	$15.39	$ 9.73
Cost/m^3 water treated	$ 2.74	$ 2.44	$ 3.08
Cost/t ore milled	$ 4.73	$ 4.21	$ 2.68

1. 720 m^3/h, 24,500 kg/d thiosalts.

Comments on the Process

The alkaline pressure oxidation process for the treatment of milling solutions has, to date, been evaluated only in laboratory experiments using synthetic solutions. The process is deemed technically uncertain for the treatment of milling effluents, and extensive experimental work would be required to define operating parameters such as retention time, pressure and temperature for the design of full-scale facilities.

The process is more complex than other oxidation methods (e.g. chlorination, H_2O_2 treatment). The personnel would be required to operate facilities generally unfamiliar to mill staff (e.g. oxygen and steam generating systems).

The process is highly energy intensive. Future operating costs would be dependent on the energy situation, and are thus unpredictable.

3.6 Reduction by Iron

Based on the observation of the disappearance of thiosalts during the wet grinding of ore using steel balls, CANMET's Mineral Sciences Laboratories have carried out laboratory tests using synthetic solutions to evaluate the destruction of thiosalts using metallic iron([1]). At 80 to 100°C, the thiosalts were found to react with iron to form FeS and H_2S. Additionally, colloids of Fe_2O_4 and FeO were formed. Based on results of the laboratory tests, a flowsheet for the reduction of thiosalts in milling effluents as shown in Figure 13 was prepared.

Mill tailings decant is passed through a settling pond for settleable solids removal. This stream and warm thickener overflows from the mill are sent to a heat exchanger for preheating using waste heat recovered from the treated effluent. Subsequently, the combined stream passes through four covered mixed tank reactors operated in series at a total retention time of two hours. Iron powder is generated on-site in a mill using iron scrap and is added to the reactors for the reduction of thiosalts. H_2S gas produced during the process is collected, and is either reacted using hydrogen peroxide or is converted to elemental sulphur using the Claus process.

The reactors are maintained at 100°C using steam, and agitators are used to keep solids in suspension to promote efficient contact between the iron and the solution. The discharge from the reactors passes through a clarifier for

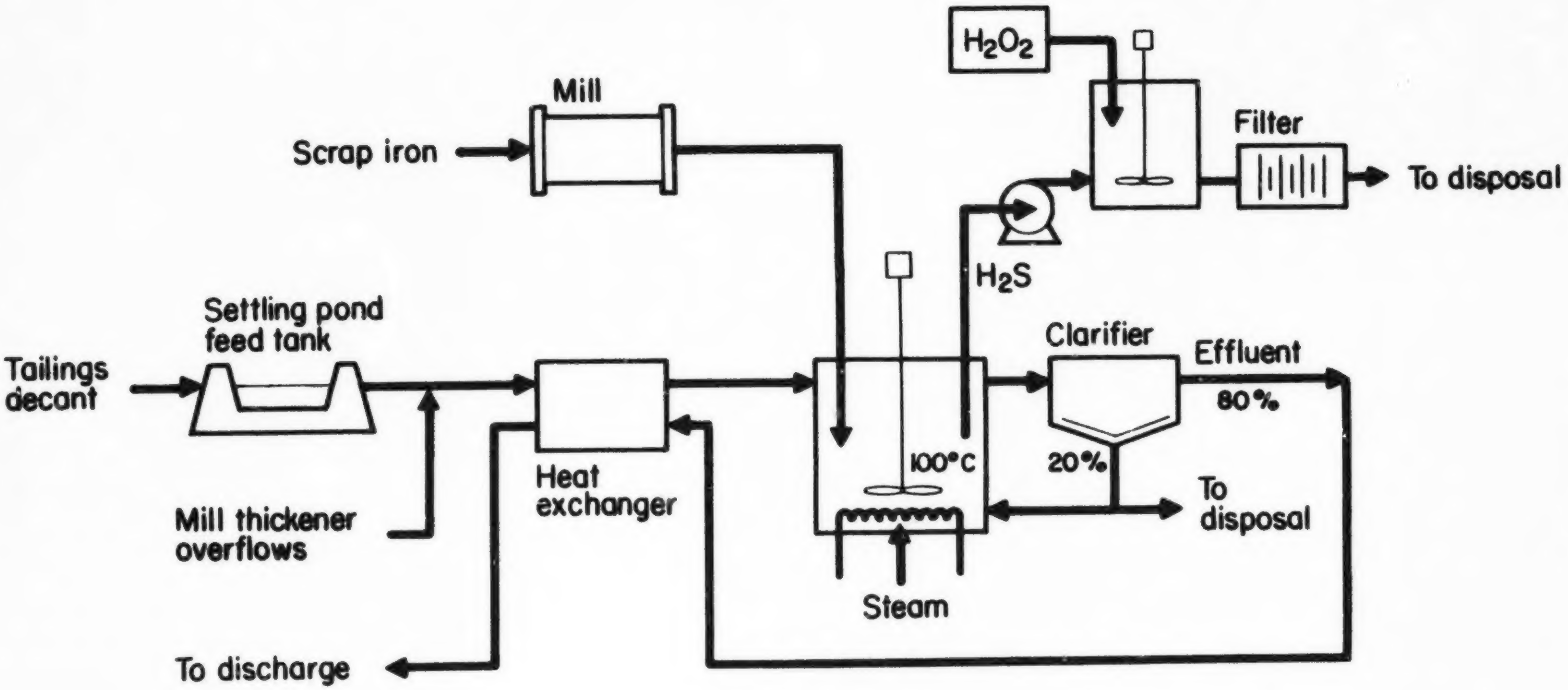

Figure 13 - Thiosalt Reduction by Iron

solid/liquid separation. Most of the clarifier underflow, which contains unreacted iron, FeS, Fe_2O_4 and FeO is recycled to the reactors. A portion of the solids is taken out continuously for disposal in the tailings ponds. The overflow at pH 8 is sent through a heat exchanger for heat recovery. The principal process design parameters used for sizing the iron reduction facilities are summarized in Table LII.

TABLE LII

IRON REDUCTION - PROCESS DESIGN PARAMETERS

Parameter	Design Basis		
	720 m^3/h. 24,500 kg/d S_2O_3	720 m^3/h. 12,300 kg/d S_2O_3	360 m^3/h. 12,300 kg/d S_2O_3
Fe to S_2O_3 weight ratio requirements	2	2	2
Total reactor retention time (h)	2	2	2
No. of reactors	4	4	4
Reactor volume(m^3)	350	350	175
Agitator power (kw/m^3)	0.33	0.33	0.33
Operating temperature (°C)	100	100	100

Cost Estimates

The capital estimates for the reduction of thiosalts with iron, followed by the reaction of the generated H_2S gas with H_2O_2, range from \$9.3 to \$13.4 million (Table LIII). The major cost items (shown in Table B-14) are heat exchangers and a steam generating plant which accounts for approximately 50% of the module cost. The use of a Claus plant to treat H_2S would increase these costs (e.g. \$3.5 million for the base case, Table LIV).

TABLE LIII - IRON REDUCTION
COSTS ESTIMATES

	Base Case	100% Flow 50% Thio	50% Flow 50% Thio
Primary Module	$'000	$'000	$'000
Bare Equipment Items	1,540.00	1,360.00	1,020.00
Misc. Unspecified Items (15% of 1)	231.00	204.00	153.00
Subtotal Bare Equipment	1,771.00	1,564.00	1,173.00
Installed Cost (Bare Equipment x 1.45)	2,567.95	2,267.80	1,700.85
Other Items - Installed Cost	0.00	0.00	0.00
Subtotal Installed Equipment	2,567.95	2,267.80	1,700.85
Services: Electrical 15%, Instr. 10%, Serv. Piping 10%, Other Serv. 5%	1,027.18	907.12	680.34
Building 3,000.00 3,000.00 2,000.00	210.00	210.00	140.00
Total Primary Module Cost (A)	3,805.13	3,384.92	2,521.19
Ht.Exch. & Steam Plant Module (B) (includes installation & services)	3,700.00	3,700.00	2,700.00
Settling Basin Cost (C)	150.00	150.00	100.00
Lime Module Cost (D)	0.00	0.00	0.00
Total Cost of Modules (A+B+C+D+E)	7,655.13	7,234.92	5,321.19
Engineering & Construction (25% of Modules)	1,913.78	1,808.73	1,330.30
Subtotal	9,568.91	9,043.65	6,651.49
Contingency (40% of Subtotal)	3,827.57	3,617.46	2,660.60
OVERALL TOTAL CAPITAL COST	13,396.48	12,661.11	9,312.08
Operating Cost	'000/y	'000/y	'000/y
Materials, Supplies, Utilities & Labour	14,119.00	11,645.00	7,060.00
Lime	0.00	0.00	0.00
Maint., Taxes & Insurance (4.5% of Tot.Cap.)	602.84	569.75	419.04
TOTAL OPERATING COST	14,721.84	12,214.75	7,479.04

TABLE LIV - IRON REDUCTION/CLAUS PROCESS
COST ESTIMATES

	Base Case
Primary Module	$'000
Bare Equipment Items	1,540.00
Misc. Unspecified Items (15% of 1)	231.00
Subtotal Bare Equipment	1,771.00
Installed Cost (Bare Equipment x 1.45)	2,567.95
Other Items - Installed Cost	0.00
Subtotal Installed Equipment	2,567.95
Services: Electrical 15%, Instr. 10%, Serv. Piping 10%, Other Serv. 5%	1,027.18
Building 3,000.00	210.00
Total Primary Module Cost (A)	3,805.13
Ht. Exch. & Steam Plant Module (B) (includes installation & services)	3,700.00
Settling Basin Cost (C)	150.00
Lime Module Cost (D)	0.00
Total Cost of Modules (A+B+C+D+E)	7,655.13
Engineering & Construction (25% of Modules)	1,913.78
Subtotal	9,568.91
Contingency (40% of Subtotal)	3,827.57
Claus Reactor (incl. 20% contingency)	3,500.00
OVERALL TOTAL CAPITAL COST	16,900.00
Operating Cost	'000/y
Materials, Supplies, Utilities & Labour	13,170.00
Lime	0.00
Maint., Taxes & Insurance (4.5% of Tot.Cap.)	602.84
TOTAL OPERATING COST	13,770.00

The operating costs, mainly for steam and iron, are highly flow dependent and range from \$7.5 to \$14.7 million per year with H_2O_2 used for H_2S destruction. Reacting the H_2S gas using the Claus process reduces the base case costs from \$14.7 to \$13.8 million per year.

The total annual costs (operating and annualized capital) are summarized in Table LV and range from \$9.3 to \$17.4 million. Treatment of H_2S gas by the Claus process or H_2O_2 oxidation results in similar costs (\$17.2 vs \$17.4 million). These expenditures translate as charges of \$2.56 to \$4.77/t ore milled, and \$2.33 to \$2.96/m^3 water treated.

Comments on the Process

The major advantage of the iron reduction process is that thiosalts are converted to sulphur or FeS, instead of SO_4. Therefore, no H_2SO_4 is generated, and there is no need for lime neutralization post-treatment, thus reducing operating and capital costs.

The reduction of thiosalts with iron has to date been evaluated only in laboratory scale experiments using synthetic solutions. Extensive development would be required to define operational parameters (e.g. retention time, temperature, iron requirements, etc.) before such a process could be seriously considered for the treatment of milling effluents.

The process is highly energy intensive, because of the need to raise up to 720 m^3/h of effluent to a temperature of 100°C. Future operating costs would be highly dependent on the

TABLE LV

REDUCTION BY IRON COST SUMMARIES

Item	Base Case[1]	100% Flow 50% S_2O_3	50% Flow 50% S_2O_3
Reduction by Iron/H_2O_2 Treatment			
Annualized total cost (millions)	$ 17.40	$ 14.75	$ 9.34
Cost/m^3 water treated	$ 2.76	$ 2.33	$ 2.96
Cost/t ore milled	$ 4.77	$ 4.04	$ 2.56
Reduction by Iron/Claus Process			
Annualized total cost (millions)	$ 17.15		
Cost/m^3 water treated	$ 2.71		
Cost/t ore milled	$ 4.69		

[1] 720 m^3/h, 24,500 kg/d thiosalts.

energy situation and are thus unpredictable.

It has been assumed in the preparation of the flowsheet that scrap iron can be ground in a mill to generate small iron particles for the process. The technical feasibility of this approach would require separate evaluation. If not feasible, specially prepared iron powder, at a much higher cost, would be required.

3.7 Activated Carbon

The removal of contaminants from water by activated carbon involves their accumulation or concentration by adsorption on the carbon surfaces. Laboratory scale tests at the Noranda Research Centre(1) have shown activated carbon to be effective for the removal of thiosalts from synthetic solutions. Because of the high cost of activated carbon, on-site carbon regeneration would be required for large-scale treatment facilities. Potential flowsheets for the treatment of milling effluents by activated carbon (based on the Noranda tests) are presented in Figures 14 and 15.

As shown, the combined milling effluent composed of mill tailings decant and thickener overflows would be sent for treatment in activated carbon columns. The granular carbon would remove not only thiosalts but also many of the other contaminants present (e.g. heavy metals). Thus, the column discharge could likely be sent to the receiving waters without additional treatment.

Two methods for the regeneration of activated carbon

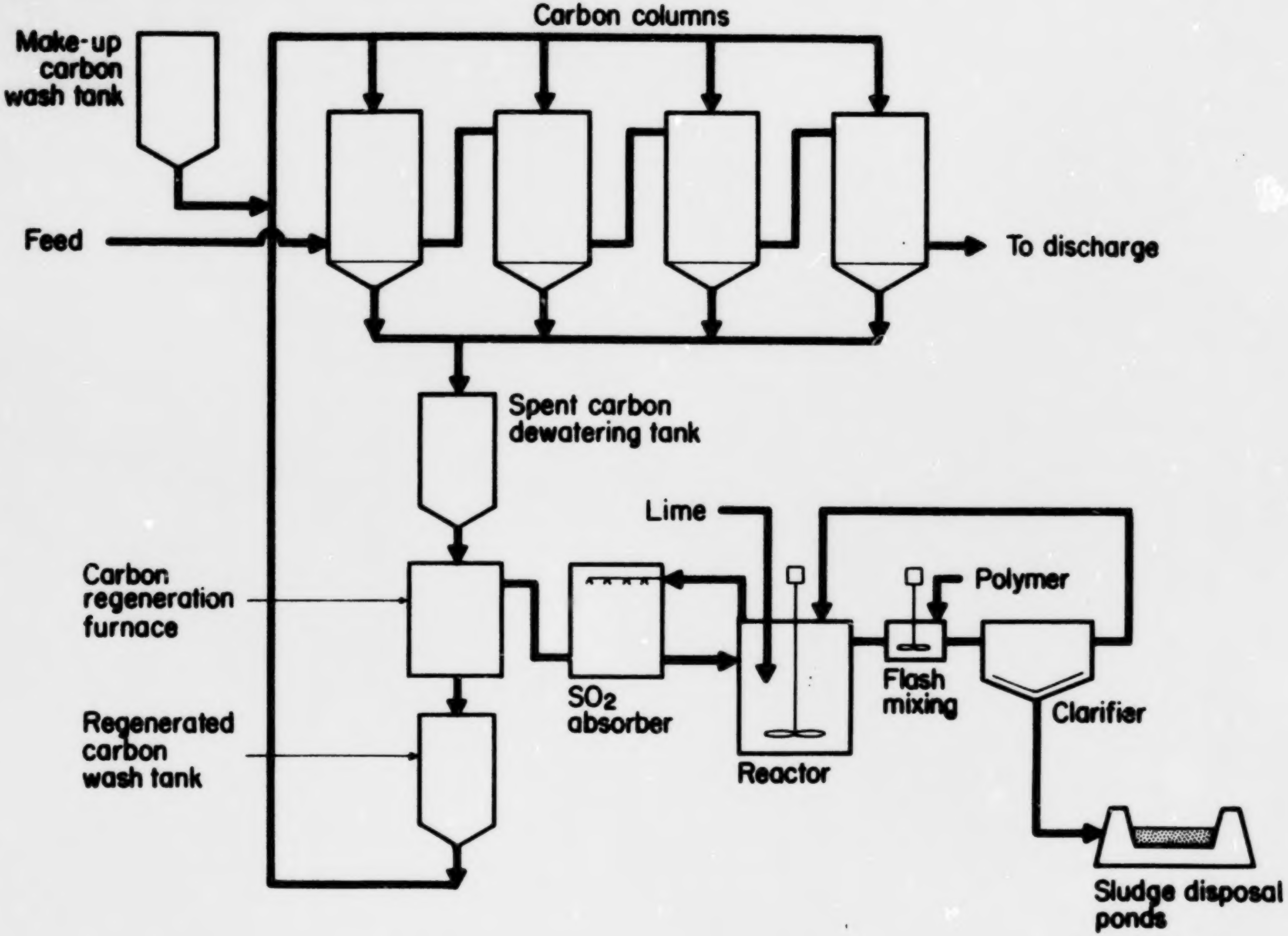

Figure 14 - Activated Carbon Adsorption with Thermal Regeneration

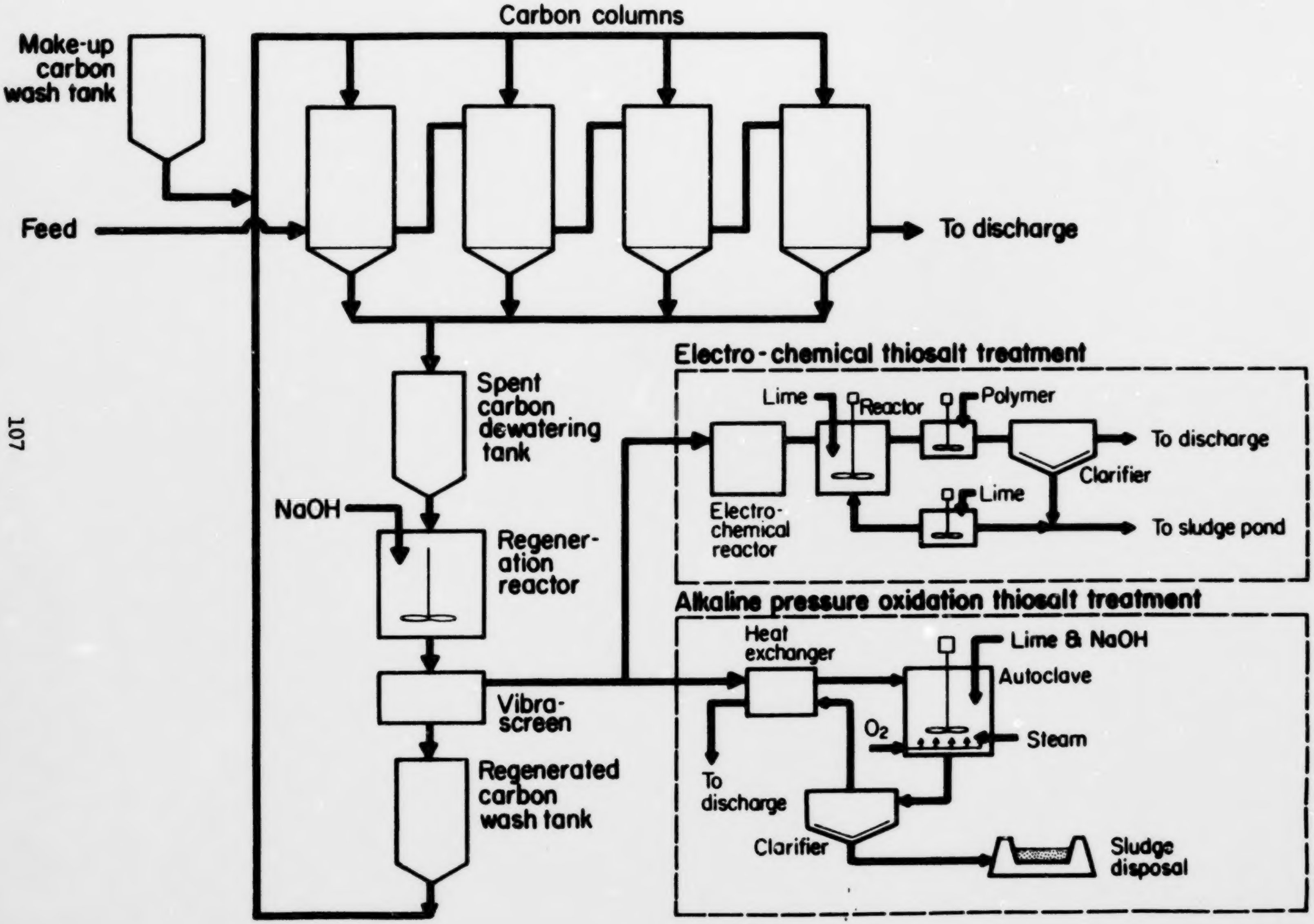

Figure 15 - Activated Carbon Adsorption with Chemical Regeneration

are generally in use: thermal and chemical. Thermal regeneration involves the use of multiple hearth furnaces (at temperatures up to 950°C) and for thiosalts, would probably occur in a three-step process as follows:

- drying;
- oxidation of thiosalts to SO_2;
- volatilization of SO_2, and other contaminants.

To prevent the release of SO_2 gas to the atmosphere, the stack gases from the multiple hearth furnace would be passed through an SO_2 scrubbing tower where the SO_2 would be converted to H_2SO_4. The resulting weak acid solution would then be neutralized in a lime treatment facility.

Sodium hydroxide solution is often used for the chemical regeneration of activated carbon. Thiosalts (and other contaminants) accumulated on the carbon surfaces would be dissolved and removed in the alkali solution which would then require treatment. Two possible methods for the destruction of thiosalts in the NaOH solutions were considered:

- Electrochemical oxidation
- Alkaline pressure oxidation

The flowsheets for these systems are described in sections 3.3 and 3.5. The principal process design parameters used for sizing the activated carbon facilities are shown in Table LVI.

TABLE LVI

ACTIVATED CARBON ADSORPTION - PROCESS DESIGN PARAMETER

Parameter	Design Basis 12,000 L/min. 24,000 kg/d S_2O_3	 12,000 L/min. 12,000 kg/d S_2O_3	 6,000 L/min. 12,000 kg/d S_2O_3
Activitated Carbon Columns			
1. Contact time (min per column)	30	15	30
2. Loading, kg carbon/kg S_2O_3	10	10	10
3. Column depth/diameter ratio	4:1	4:1	4:1
4. Column volume, m^3 each	200	200	100
Thermal Regeneration System			
1. Total loading, t/day	270	135	135
2. Multiple hearth furnace sizing, kg carbon/m^2 area	1.7	1.7	1.7
3. Total furnace surface, m^2	200	100	100
4. Steam requirements, kg/kg carbon	1	1	1
5. Heat requirements kg/kg carbon	1600	1600	1600
6. Regeneration loss of carbon, %	10	10	10
SO_2 Scrubber System			
1. Gas input, kg/h.	28500	14300	14300
2. SO_2 input, kg/h.	1200	600	600
Alkaline Regeneration System			
1. NaOH solution strength (normality)	1	1	1
2. Thiosalt loading, g/L	50	50	50
3. Caustic requirements, t/day	22	11	11
4. Regeneration loss of carbon, %	1	1	1

Cost Estimates

Costs were derived for activated carbon treatment followed by three methods of carbon regeneration (Tables LVII to LIX). Lowest capital costs are associated with chemical regeneration; alkaline oxidation of waste solutions result in lower costs than electrooxidation ($12.7 to $19 million versus $15.2 to $23.6 million). Thermal regeneration of the activated carbon results in substantially higher capital charges ($24 to $37 million). In all cases, the major portion of the costs is for the regeneration facilities (Tables B15 to B17, Appendix B).

Operating costs for thermal regeneration are very high ($33 to $65 million/y) mainly due to furnace operation and carbon make-up charges. Chemical regeneration annual operating costs amount to $4.3 to $7.8 million for alkaline oxidation, and $4.9 to $9.1 million for electrooxidation.

The total annual costs (operating and annualized capital) are summarized in Table LX. Lowest costs occur for carbon treatment with chemical regeneration using caustic followed by alkaline oxidation, and total $6.8 to $11.7 million/y. This corresponds to $1.86 to $3.20/t ore milled, and $1.26 to $2.15/m^3 water treated.

Comments on the Process

Few practical data are available regarding the treatment of thiosalt-containing milling effluents with activated carbon. Available results are only from preliminary bench-scale studies conducted with synthetic effluents. Treatment of actual

TABLE LVII - ACTIVATED CARBON/THERMAL REGENERATION COST ESTIMATES

	Base Case	100% Flow 50% Thio	50% Flow 50% Thio
Primary Module	$'000	$'000	$'000
Bare Equipment Items	6,265.00	4,100.00	4,100.00
Misc. Unspecified Items (15% of 1)	939.75	615.00	615.00
Subtotal Bare Equipment	7,204.75	4,715.00	4,715.00
Installed Cost (Bare Equipment x 1.45)	10,446.89	6,836.75	6,836.75
Other Items - Installed Cost	3,950.00	3,100.00	2,600.00
Subtotal Installed Equipment	14,396.89	9,936.75	9,436.75
Services: Electrical 15%, Instr. 10%, Serv. Piping 10%, Other Serv. 5%	5,758.76	3,974.70	3,774.70
Building 4,000.00 3,000.00 3,000.00	280.00	210.00	210.00
Total Primary Module Cost (A)	20,435.64	14,121.45	13,421.45
Settling Basin Cost (B)	150.00	150.00	100.00
Lime Module Cost (C)	0.00	0.00	0.00
Total Cost of Modules (A + B + C)	20,585.64	14,271.45	13,521.45
Engineering & Construction (25% of A+B+C)	5,146.41	3,567.86	3,380.36
Subtotal	25,732.05	17,839.31	16,901.81
Contingency (40% of Subtotal)	10,292.82	7,135.73	6,760.73
Initial Catalyst Charge (Incl. 20% Conting.)	936.00	936.00	468.00
OVERALL TOTAL CAPITAL COST	36,960.87	25,911.04	24,130.54
Operating Cost	'000/y	'000/y	'000/y
Materials, Supplies, Utilities & Labour	63,520.00	31,960.00	31,960.00
Lime	0.00	0.00	0.00
Maint., Taxes & Insurance (4.5% of Tot.Cap.)	1,663.24	1,166.00	1,085.87
TOTAL OPERATING COST	65,183.24	33,126.00	33,045.87

TABLE LVIII - ACTIVATED CARBON/ELECTROCHEMICAL OXIDATION - COST ESTIMATES

	Base Case	100% Flow 50% Thio	50% Flow 50% Thio
Primary Module	$'000	$'000	$'000
Bare Equipment Items	175.00	120.00	115.00
Misc. Unspectified Items (15% of 1)	26.25	18.00	17.25
Subtotal Bare Equipment	201.15	138.00	132.25
Installed Cost (Bare Equipment x 1.45)	291.81	200.10	191.76
Other Items - Installed Cost	1,950.00	1,800.00	1,300.00
Subtotal Installed Equipment	2,241.81	2,000.10	1,491.76
Services: Electrical 15%, Instr. 10%, Serv. Piping 10%, Other Serv. 5%	896.73	800.04	596.71
Building 2,000.00 1,750.00 1,750.00	140.00	122.50	122.50
Total Primary Module Cost (A)	3,278.54	2,922.64	2,210.97
ElectroChem. Module (incl. bldg & services) (B)	8,000.00	5,000.00	5,000.00
Settling Basin Cost (C)	150.00	150.00	100.00
Lime Module Cost (D)	1,520.00	1,100.00	1,100.00
Total Cost of Modules (A+B+C+D)	12,948.54	9,172.64	8,410.97
Engineering & Construction (25% of Modules)	3,237.13	2,293.16	2,102.74
Subtotal	16,185.67	11,465.80	10,513.71
Contingency (40% of Subtotal)	6,474.27	4,586.32	4,205.48
Initial Catalyst Charge (Incl. 20% Conting.)	936.00	936.00	468.00
OVERALL TOTAL CAPITAL COST	23,595.94	16,988.12	15,187.19
Operating Cost	'000/y.	'000/y.	'000/y.
Materials, Supplies, Utilities & Labour	7,700.00	4,900.00	4,100.00
Lime	300.00	150.00	150.00
Maint., Taxes & Insurance (4.5% of Tot.Cap.)	1,061.82	764.47	683.42
TOTAL OPERATING COST	9,061.82	5,814.47	4,933.43

TABLE LIX - ACTIVATED CARBON/ALKALINE PRESSURE OXIDATION - COST ESTIMATES

	Base Case	100% Flow 50% Thio	50% Flow 50% Thio
Primary Module	$'000	$'000	$'000
Bare Equipment Items	175.00	120.00	115.00
Misc. Unspectified Items (15% of 1)	26.25	18.00	17.25
Subtotal Bare Equipment	201.25	138.00	132.25
Installed Cost (Bare Equipment x 1.45)	291.81	200.10	191.76
Other Items - Installed Cost	1,950.00	1,800.00	1,300.00
Subtotal Installed Equipment	2,241.81	2,000.10	1,491.76
Services: Electrical 15%, Instr. 10%, Serv. Piping 10%, Other Serv. 5%	896.73	800.04	596.71
Building 2,000.00 1,750.00 1,750.00	140.00	122.50	122.50
Total Primary Module Cost (A)	3,278.54	2,922.64	2,210.97
Alk. Oxid. Module (incl. bldg & services) (B)	1,800.00	1,200.00	1,200.00
Oxygen Plant (includes bldg & services) (C)	3,600.00	2,400.00	2,400.00
Settling Basin Cost (D)	150.00	150.00	100.00
Lime Module Cost (E)	1,520.00	1,100.00	1,100.00
Total Cost of Modules (A+B+C+D+E)	10,348.54	7,772.64	7,010.97
Engineering & Construction (25% of Modules)	2,587.13	1,943.16	1,752.74
Subtotal	12,935.67	9,715.80	8,763.71
Contingency (40% of Subtotal)	5,174.27	3,886.32	3,505.48
Initial Catalyst Charge (Incl. 20% Conting.)	936.00	936.00	468.00
OVERALL TOTAL CAPITAL COST	19,045.94	14,538.12	12,737.19
Operating Cost	'000/y	'000/y	'000/y
Materials, Supplies, Utilities & Labour	6,687.00	4,194.00	3,544.00
Lime	300.00	150.00	150.00
Maint., Taxes & Insurance (4.5% of Tot.Cap.)	857.07	654.22	573.17
TOTAL OPERATING COST	7,844.07	4,998.22	4,267.17

TABLE LX

ACTIVATED CARBON COST SUMMARIES

Item	Base Case[1.]	100% Flow 50% S_2O_3	50% Flow 50% S_2O_3
Activated Carbon/Thermal Regeneration			
Annualized total cost (millions)	\$ 72.58	\$ 38.31	\$ 38.87
Cost/m^3 water treated	\$ 11.51	\$ 6.07	\$ 12.30
Cost/t ore milled	\$ 19.91	\$ 10.49	\$ 10.65
Activated Carbon/ Electrooxidation			
Annualized total cost (millions)	\$ 13.78	\$ 9.21	\$ 7.97
Cost/m^3 water treated	\$ 2.18	\$ 1.46	\$ 2.50
Cost/t ore milled	\$ 3.77	\$ 2.51	\$ 2.16
Activated Carbon/ Alkaline Oxidation			
Annualized total cost (millions)	\$ 11.67	\$ 7.91	\$ 6.81
Cost/m^3 water treated	\$ 1.84	\$ 1.26	\$ 2.15
Cost/t ore milled	\$ 3.20	\$ 2.17	\$ 1.86

1. 720 m^3/h, 24,500 kg/d thiosalts.

effluents would entail a great number of unknowns requiring lengthy testwork to clarify. Such factors include:

- Carbon requirements may be higher than assumed because of adsorption of other contaminants (e.g. metals, SO_4) which would certainly reduce the available surface for thiosalt removal.

- Scaling on carbon surfaces which would reduce carbon efficiencies, and make regeneration more difficult.

- The technical aspects of thermal regeneration are highly uncertain. Sodium thiosulphate decomposes at 233°C, as follows([23]):

$$4Na_2S_2O_3 \rightarrow 3Na_2SO_4 + Na_2S_5$$

Further heating to 440 to 500°C results in the generation of Na_2S and SO_2. Na_2SO_4 is decomposed, especially in the presence of impurities, at >1100°C to SO_2, SO_3, O_2. In the presence of carbon at 1000°C, Na_2S or SO_2 are formed, depending on the presence of air.

The regeneration of carbon with caustic, followed by either electrochemical treatment or alkaline pressure oxidation is untested, and would require extensive technical evaluation.

The treatment of milling effluents by activated carbon would remove, besides thiosalts, most of the other contaminants present (e.g. metals, sulphate), and would thus produce high quality water for discharge. This water could be recycled for use in the mill, and thus may reduce mill operating costs.

3.8 Disposal to the Sea

Disposal of effluents to the sea is widely practised by industry both in North America and in Europe. It generally involves control over discharge paths to ensure adequate dilution and negligible changes in seawater concentrations. The mine and mill of BMS are located approximately 30 km from Bathurst Harbour and the Baie des Chaleurs. Piping of the milling effluents to the sea was evaluated as one of the options for thiosalt disposal. As indicated in Figure 16, the system involves pre-treatment of the combined tailings decant and mill thickener overflows by lime neutralization to pH 9 for the removal of dissolved metals. This would take place in a conventional lime neutralization plant (consisting of reactors, polymer addition, and a clarifier) operated in the high sludge density mode. The neutralized effluent would be collected in a sump, and would then be pumped through a polyethylene pipe. Four routes were considered; the route selected was the shortest and, as shown in Figure 17, follows along the South Little River, the Middle River highway, around the Town of Bathurst, and into Nepisiguit Bay. A 1.5 km submerged diffuser is used for the discharge of the effluent.

Cost Estimates

The capital cost estimates for the disposal of milling effluents to the sea are presented in Table LXI. Major cost items include the lime neutralization plant, piping, and pipe installation and burial (Table B-18, Appendix B). Parts of the land used for the pipeline route currently are owned by the

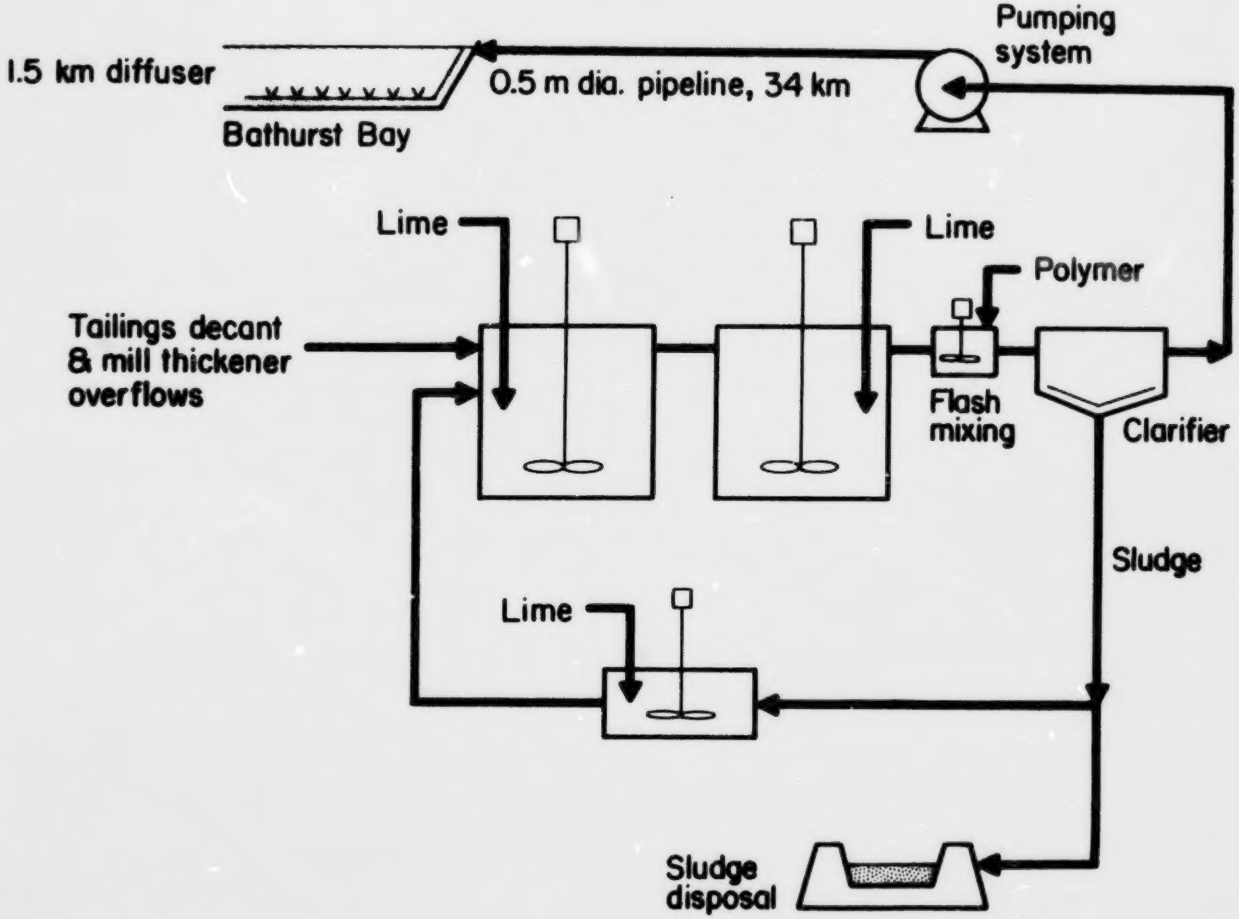

Figure 16 - Disposal to the Sea

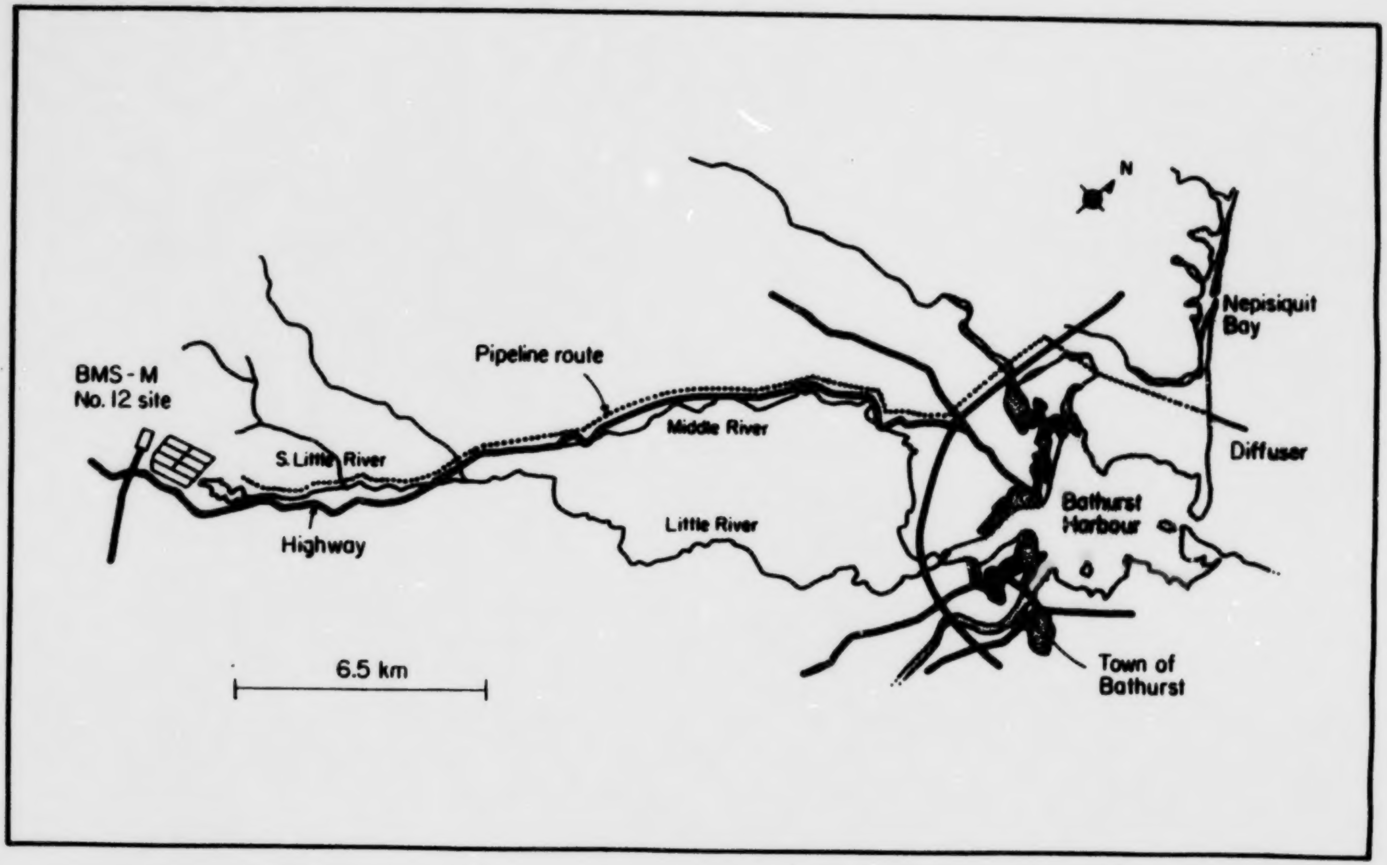

Figure 17 - Suggested Pipeline Route

TABLE LXI - DISPOSAL TO THE SEA - COST ESTIMATE

	Base Case	100% Flow 50% Thio	50% Flow 50% Thio
Primary Module	$'000	$'000	$'000
Bare Equipment Items	355.00	355.00	250.00
Misc. Unspectified Items (15% of 1)	53.25	53.25	37.50
Subtotal Bare Equipment	408.25	408.25	287.50
Installed Cost (Bare Equipment x 1.45)	591.96	591.96	416.88
Other Items - Installed Cost	0.00	0.00	0.00
Subtotal Installed Equipment	591.96	591.96	416.88
Services: Electrical 15%, Instr. 10%, Serv. Piping 10%, Other Serv. 5%	236.79	236.79	166.75
Building 600.00 600.00 500.00	42.00	42.00	35.00
Total Primary Module Cost (A)	870.75	870.75	618.63
Pipeline and Diffuser (installed) (B)	11,160.00	11,160.00	9,460.00
Settling Basin Cost (C)	150.00	150.00	100.00
Lime Module Cost (D)	2,930.00	2,800.00	2,400.00
Total Cost of Modules (A+B+C+D)	15,110.75	14,980.75	12,578.63
Engineering & Construction (25% of Modules)	3,777.69	3,745.19	3,144.66
Subtotal	18,888.43	18,725.93	15,723.28
Contingency (40% of Subtotal)	7,555.37	7,490.37	6,289.31
Cost of Right of Way	2,500.00	2,500.00	2,500.00
OVERALL TOTAL CAPITAL COST	28,943.81	28,716.31	24,512.59
Operating Cost	'000/y	'000/y	'000/y
Materials, Supplies, Utilities & Labour	70.00	70.00	35.00
Lime	183.00	92.00	92.00
Maint., Taxes & Insurance (4.5% of Tot.Cap. exc. Rt. of Way)	1,189.97	1,179.73	990.57
TOTAL OPERATING COST	1,442.97	1,341.73	1,117.57

Consolidated Bathurst Company and the Government of New Brunswick. It has been assumed that a right-of-way costing $2.5 million would be required. Overall capital costs are similar for the three systems, and range from $24.5 to $28.9 million.

Operating costs, mainly for maintenance, taxes, and insurance, range from $1.1 to $1.9 million/per year. The total annual costs (operating and annualized capital) are summarized in Table LXII, and range from $6.0 to $7.2 million. This corresponds to $1.65 to $1.98/t ore milled, and $1.11 to $1.90/m^3 water treated.

TABLE LXII
DISPOSAL TO THE SEA COST SUMMARIES

Item	Base Case[1]	100% Flow 50% S_2O_3	50% Flow 50% S_2O_3
Annualized total cost (millions)	$ 7.23	$ 7.08	$ 6.02
Cost/m^3 water treated	$ 1.13	$ 1.11	$ 1.90
Cost/t ore milled	$ 1.98	$ 1.93	$ 1.65

1. 720 m^3/h, 24,300 kg/d thiosalts.

Comments on the Process

The disposal of effluents through a pipeline to the sea, although simple in concept, entails a variety of potential problems:

1. A pipeline leak would result in the discharge of thiosalts to the environment and the possible generation on acid between the plant and the sea.

2. Flow of lime-treated effluents through a pipeline will result in scaling. It is not known whether routine maintenance

or the use of scale inhibiting reagents will be sufficient to control this problem.

3. There may be social aspects involved, with opposition arising from the general population against the disposal of relatively untreated effluents to the sea.

4. A pipeline alternative would probably require lengthy and costly environmental impact studies and hearings which may result in an unfavourable verdict. The potential costs of such studies have not been considered in the current estimates.

3.9 Natural Degradation

One of the first thiosalt treatment schemes in use was a long retention time pond allowing natural degradation (likely in the presence of bacteria) of thiosalts to take place. The pond put into operation at BMS in 1970 has however achieved limited thiosalt degradation (generally around 50%/y average) due to such factors as a short retention time (25 days) and probable short-circuiting.

A natural degradation facility with a larger retention time (say one to two years) would probably achieve more successful results. For the purposes of this study, preliminary costs were derived for a relatively shallow (3 m depth) system made up of three ponds in series, followed by conventional lime treatment facilities for the neutralization of generated acid (Figure 18). Estimated costs for such a system are indicated in Table LXIII. As shown, capital costs range from $9.4 to $12.4 million, with operating costs estimated at $0.97 to $1.61

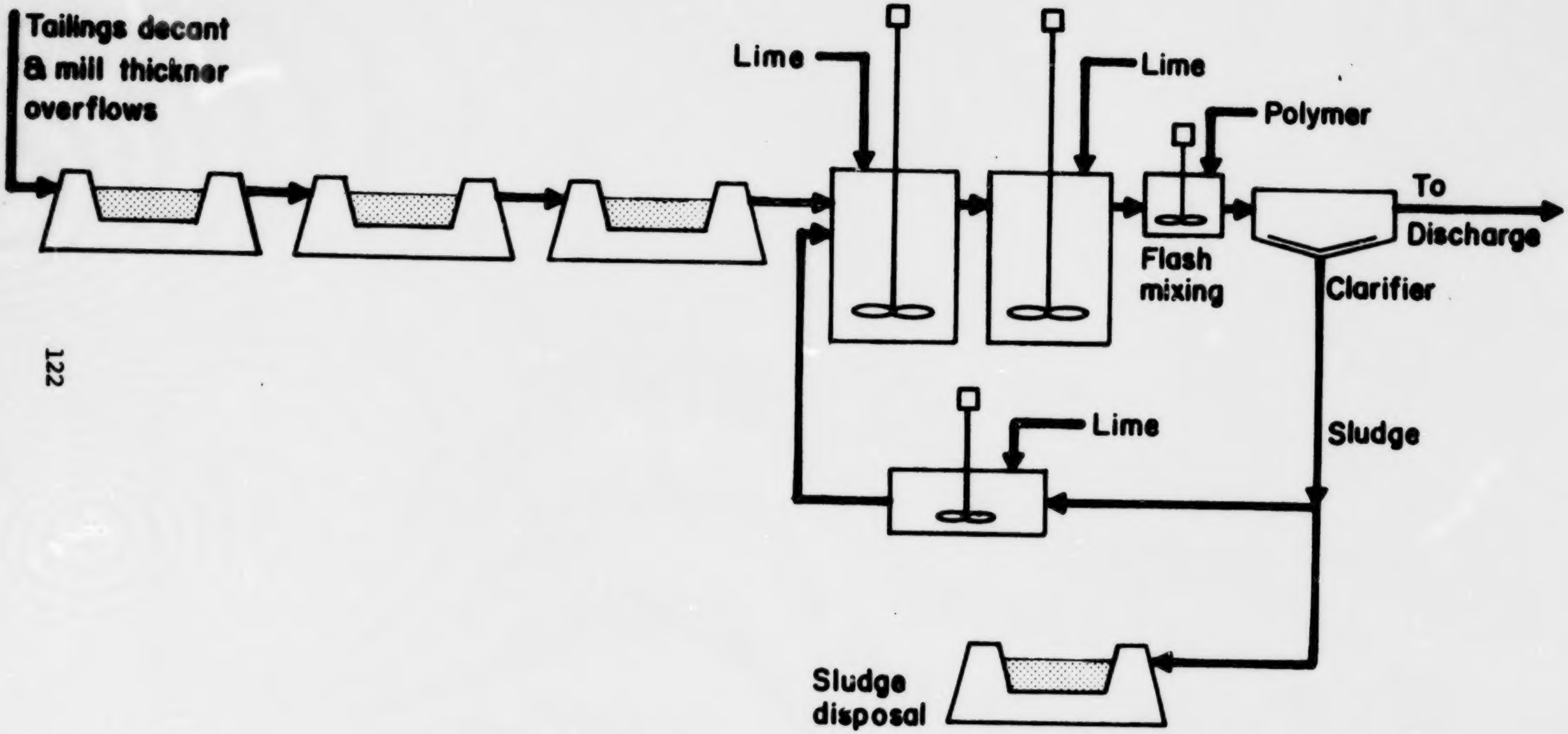

Figure 18 - Natural Degradation

TABLE LXIII

NATURAL DEGRADATION - COST ESTIMATES

Item	Base Case	100% Flow 50% S_2O_3	50% Flow 50% S_2O_3
	$,000	$,000	$,000
Natural Degradation Ponds (A)	$ 3,500	$ 3,500	$ 2,500
Lime Module (B)	$ 3,580	$ 3,240	$ 2,850
Total Cost	$ 7,080	$ 6,740	$ 5,350
Engineering and Construction (25% of A + B)	$ 1,770	$ 1,690	$ 1,338
Sub Total	$ 8,850	$ 8,430	$ 6,688
Contingency (40% of Sub Total)	$ 3,540	$ 3,370	$ 2,675
OVERALL TOTAL CAPITAL COST	$12,390	$11,800	$ 9,393
Operating Cost	$,000/y	$,000/y	$,000/y
Labour	$ 50	$ 50	$ 50
Lime	$ 1,000	$ 500	$ 500
Maint., Taxes & Insurance	$ 558	$ 531	$ 421
TOTAL OPERATING COST	$ 1,608	$ 1,081	$ 971

million/y. Total annualized costs, shown in Table LXIV vary from \$2.9 to \$4.1 million per year. This amounts to \$0.80 to \$1.12/t ore milled, or \$0.53 to \$0.91/m^3 water treated.

The technical feasibility of this approach is uncertain, and pilot scale trials would be required to determine whether complete degradation would occur. The land requirements for such a pond system are quite extensive and may not be available everywhere.

TABLE LXIV

NATURAL DEGRADATION - COST SUMMARIES

Item	Base Case[1]	100% Flow 50% S_2O_3	50% Flow 50% S_2O_3
Annualized total cost (millions)	\$ 4.10	\$ 3.40	\$ 2.90
Cost/m^2 water treated	\$ 0.64	\$ 0.53	\$ 0.91
Cost/t ore milled	\$ 1.12	\$ 0.92	\$ 0.80

[1] 720 m^3/h, 24,300 kg/d thiosalts.

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APPENDIX A

RESULTS OF BIOLOGICAL AIR OXIDATION PILOT STUDY

SUMMARY

Pilot plant studies to evaluate the biological treatment of thiosalts in milling effluents have been conducted at Brunswick Mining and Smelting (BMS) between 1979 and 1984. Tests with thickener overflow streams have shown inconsistent performance due to periodic high solids discharge from the thickeners, high lead concentrations, and wide variability of the thiosalt concentrations in the effluents. Tests with the tailings decant stream were more successful, with complete thiosalt removal being achieved at temperatures as low as 1°C.

1. THICKENER OVERFLOW TESTS

1.1 1979 to mid-1981 Period

Engineering, design and construction of the thiosalt pilot plant was carried out between June and November 1979. A flowsheet and layout drawing of the facility, which consisted of a 250,000-Imp. gal. settling pond, a 55,000 Imp. gal. three-compartment Aerated Pond (AP), a 29,000-Imp. gal. void volume Aerated Pond (ARP) as well as three packed aerated columns, are presented in Figures A-1, A-2.

Pilot plant operation was initiated in the static mode in early December 1979 by filling both the ARP and AP with BMS biostabilization pond effluent which contains naturally-occurring colonies of thiosalt oxidizing bacteria. Phosphate and nitrate nutrients were added as phosphoric acid and ammonium sulphate in a ratio of S_2O_3:NH_4:PO_4=1000:8:5 and air was blown into the ponds

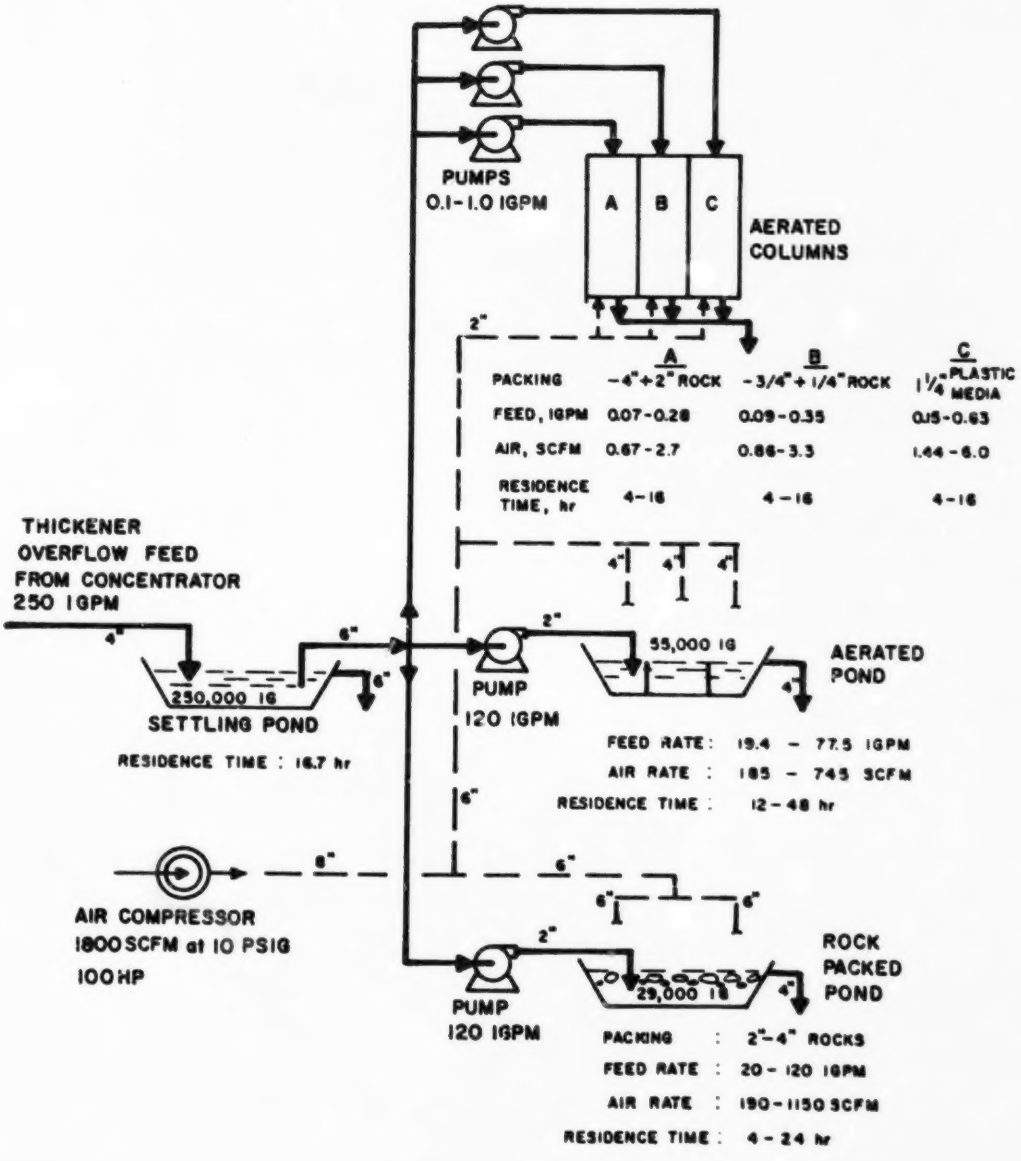

Figure A-1 - Flowsheet of the Noranda Research Thiosalt Pilot Plant

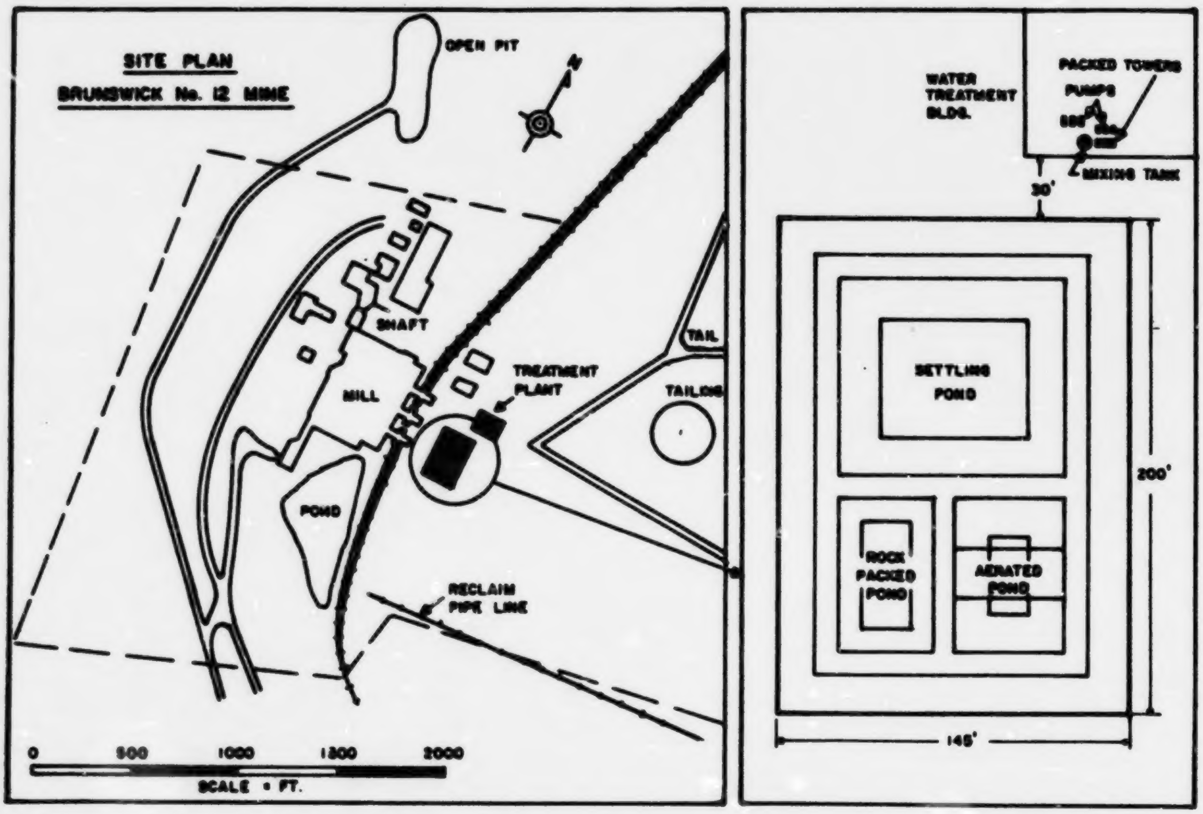

Figure A-2 - Schematic Diagram of the Noranda Research Thiosalt Pilot-Plant Layout using the air distribution system. Due to cold temperatures, all attempts at activating the bacterial process in both the ARP and the AP between December 1979 and March 1980 proved futile with minimal decreases in pH and thiosalt concentrations being observed despite such measures as reduction in residence time to reduce heat loss. Biological activity increased with warmer ambient temperatures during April 1980, resulting in reduced thiosalt and pH levels. Unfortunately, strike action at BMS in May 1980 halted further tests.

At the end of the BMS strike in the fall of 1980, pilot tests were resumed using only the aerated rock pond to concentrate the test effort. As during the previous year, cold weather hampered proper bacterial acclimatization and ARP liquor temperatures were down to 0 to 3°C in early November 1980. To prevent the previous year's undesirable occurrences, a steam header was installed at the pilot plant and stream was piped into the ARP, thus raising the pond liquor temperature to 20°C. This, in turn, promoted bacterial activity. Continuous feed forward of combined mill thickener streams to the ARP started in mid-November and continued to the end of February 1981. After acclimatization, the warm feed temperature allowed the pond to operate with no additional inputs of steam. Although high thiosalt removals were achieved (up to 80-93% with feed thiosalt levels at 1500 to 3300 mg/L as S_2O_3), numerous operating problems were encountered which prevented extended run duration. These included the malfunction of the mill polymer system resulting in poor thickener operation and high concentrate loadings in the thickener overflows; freeze-up of the thickener take-off and ARP overflow lines; and discharge of spent lime with the thickener overflows. Corrective actions by BMS included the installation of a separate line for the discharge of spent lime slurry to the tailings area, repairs to the polymer feed system, and thawing and/or replacement of frozen lines. By February 1981, the pilot tests were halted to allow cleanup of the solids in the settling pond.

1.2 Mid 1981 to 1982 Period

Operation of the ARP was resumed on 15 May 1981 in the static mode, followed a few days later by partial feed-forward (6h/day). Continuous operation was started on 24 June 1981 and was maintained more or less without interruption for a 30-week period to the end of January 1982. Then a variety of problems caused disruptions of pilot plant operations to the end of August 1982, and these precluded extended runs. Subsequently, operations resumed, and the ARP was operated almost continuously during 17 weeks to mid-December 1982 when the decision was taken to halt further tests.

Tests using two pilot columns were initiated in September 1981 following a five-week period of aeration in the static mode (the third column leaked and was not used) and continued to the end of January 1982 when they were halted by line freeze-up. No further tests were conducted.

Limited tests were conducted with the Aerated Pond, with roughly 10 weeks of operation occuring between October 1982 and mid-December 1982, when the pilot tests were halted.

1.3 Results

The results of the Aerated Rock Pond tests are presented in Figure A-3 for the July to December 1981 period. Based on these results, the following can be concluded:

a) Although complete thiosalt removals have been attained periodically, the pilot tests have not demonstrated that thiosalt concentrations can be consistently reduced to discharge levels of 100 to 200 mg/L. One of the major factors for inconsistent pilot plant performance has been the poor operation of the mill thickeners. This has resulted in high concentrate solids discharges in the thickener overflows, elevated metal levels (particularly Pb, see below) in the feed to the pilot plant and the frequent need to clean out the pilot settling pond.

b) Thiosalt removal appeared to be related to the lead levels in the ARP effluent; poor performance was observed when the lead concentrations were above 10 to 15 mg/L. Better performance (80% removal) was achieved with Pb levels below 10 mg/L and residence times of 24 hours or more.

c) There are indications that sudden increases in feed thiosalt concentrations may result in reduced performance, indicating susceptibility of the process to shock thiosalt inputs.

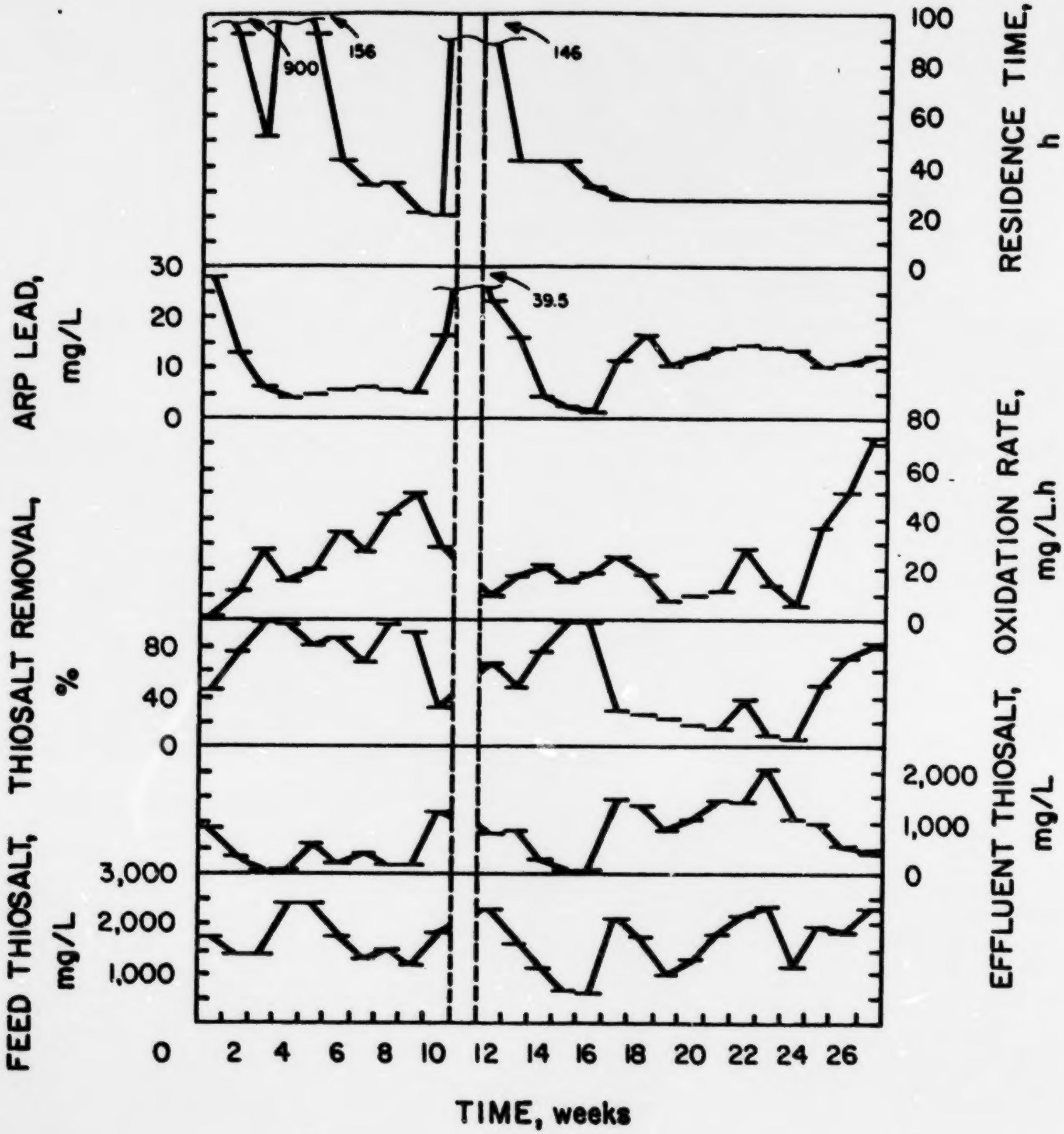

Figure A-3 - Aerated Rock-Filled Pond - Weekly Average Data, July-December 1981

The Aerated Pond was operated for a 10- week period during the final part of the study, between September and December 1982. Thiosalt removals ranged from 13 to 61%, and thiosalt concentrations in the final effluent were from 325 to 1550 mg/L. Pond effluent temperatures ranged from 7 to 15°C and were thus below the optimal temperatures of 25 to 30°C, suggesting that adequate acclimatization was not achieved and that pond performance was below optimal levels.

In general, the performance of the packed aerated columns followed the pattern observed in the ARP tests, except that the oxidation rates were considerably higher (up to 160 mg/L.h versus 70 mg/L.h for the ARP), probably due to differences in geometry (i.e. no sloped walls) and better air mixing.

2. TAILINGS DECANT TESTS

Pilot tests were conducted during 1983-84 with BMS tailings decant. A summary of the Aerated Rock Pond (ARP) results for the period January to December 1984 is presented in Figure A-4, and similar results for the Aerated Pond (AP) for the period June 1983 to June 1984 are presented in Figure A-5. Based on the data, the following is concluded:

1. Packing or surface material for bacterial growth is essential for effective biological treatment of thiosalts. Complete thiosalt removal was achieved in the pilot Aerated Rock

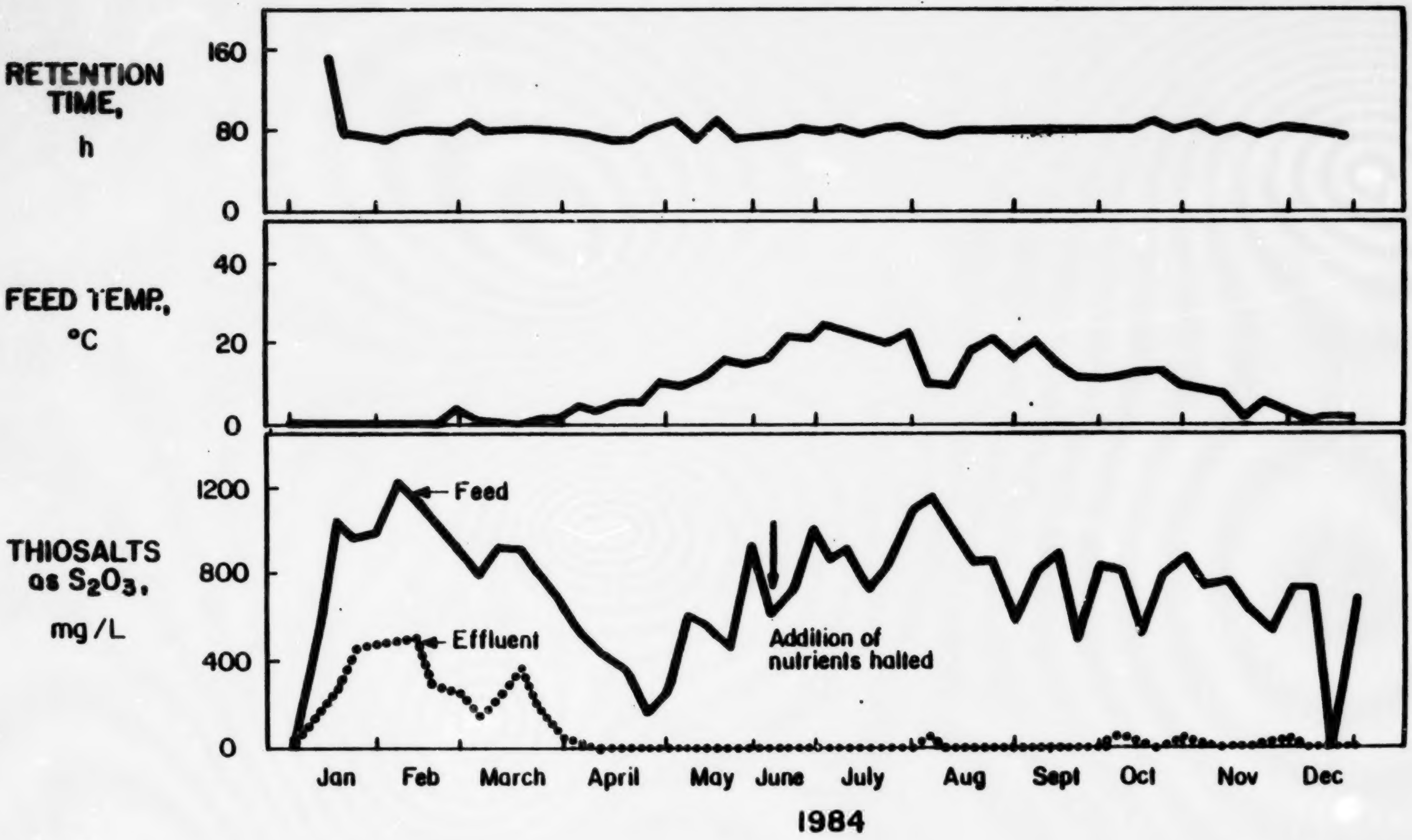

Figure A-4 - Aerated Rock Pond - 1984 Pilot Test Results

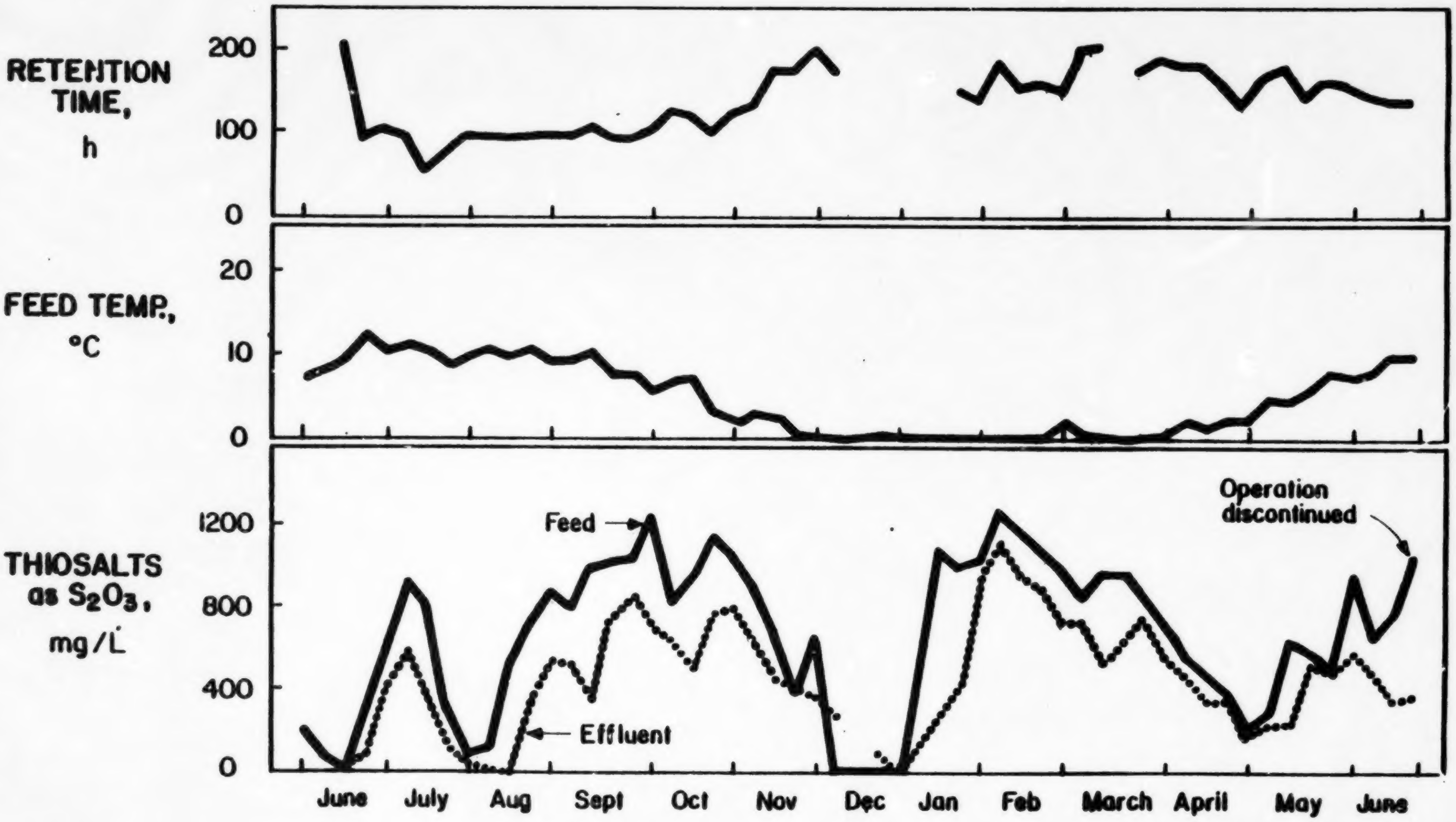

Figure A-5 - Aerated Pond - Pilot Test Results June 1983 - June 1984

Pond at temperatures as low as 1°C. The pilot Aerated Pond performed poorly, with only 60% of the thiosalts being destroyed at 20°C.

2. A retention time of 65 to 80 h was needed for the treatment of BMS-M tailings decant in the pilot ARP.

3. Cessation of nutrient addition to the ARP did not noticeably affect performance (Figure 4). This is probably due to the presence of phosphates and ammonia in BMS-M effluents as noted in previous studies.

4. Lead concentrations in the feed and effluent streams were generally below 4 mg/L, and had no apparent effect on thiosalt removal efficiencies.

APPENDIX B

DETAILED CAPITAL AND OPERATING COSTS

APPENDIX B - CAPITAL AND OPERATING COSTS BREAK-DOWN[1]

Table B-1 - Biological Air Oxidation
Table B-2 - Biological Air Oxidation (with H_2O_2 Post-Treatment)
Table B-3 - PVP-Cu Catalyzed Air Oxidation
Table B-4 - Copper Sulphide Catalyzed Air Oxidation
Table B-5 - Hydrogen Peroxide Oxidation
Table B-6 - Chlorine Oxidation
Table B-7 - Ozone Oxidation
Table B-8 - SO_2 - Air Oxidation
Table B-9 - Electrochemical Oxidation
Table B-10 - Reverse Osmosis and Evaporation
Table B-11 - Reverse Osmosis and Alkaline Pressure Oxidation
Table B-12 - Reverse Osmosis and Electrochemical Oxidation
Table B-13 - Alkaline Pressure Oxidation
Table B-14 - Reduction by Iron
Table B-15 - Activated Carbon and Thermal Regeneration
Table B-16 - Activated Carbon and Electrochemical Oxidation
Table B-17 - Activated Carbon and Alkaline Pressure Oxidation
Table B-18 - Disposal to the Sea

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[1] Base Case - 720 m^3/h, 24,300 kg/d thiosalts.

TABLE B-1

BIOLOGICAL AIR OXIDATION

CAPITAL COSTS

No.	Item Description	Quantity/Size	Unit Cost	Total Cost
1	Dam construction (cut and fill)	100,000 m^3	$5.20/$m^3$	$ 520,000
2.	Overflow weirs, concrete	8	$5,000	$ 50,000
3.	Washed rock, 5 to 10 cm size	115,000 m^3	$13/$m^3$	$1,500,000
4.	Air piping, schedule 40, steel	0.6 m dia, 610 m	$13/m	$ 80,000
5.	Air piping, schedule 40, PVC	0.13 m dia, 11,700 m	$4/m	$ 461,000
6.	Blowers	4-450 m^3/min	$125,000	$ 500,000
7.	Nutrient storage facilities			$ 14,000
8.	Nutrient make-up and feed system			$ 15,000
9.	Building	75 m^2	$750/$m^2$	$ 56,000

OPERATING COSTS

No.	Item Description	Amount	Unit Cost	Total Cost
1.	Electric power	14.5 x 10^6 kW h	$0.035/kW h	$ 506,000
2.	Nutrients - Diammonium phosphate	70 t/y	$1700/t	$ 119,000
	- Ammonium sulphate	250 t/y	$370/t	$ 92,500
3.	Operating labour	4 my	$50,000/my	$ 200,000

TABLE B-2

BIOLOGICAL AIR OXIDATION (WITH H_2O_2 POST-TREATMENT)

CAPITAL COSTS

No.	Item Description	Quantity/Size	Unit Cost	Total Cost
1.	Dam construction (cut and fill)	100,000 m^3	\$5.20	\$ 520,000
2.	Overflow weirs, concrete	8	\$5000	\$ 40,000
3.	Washed rock, 5 to 10 cm size	115,000 m^3	\$13/m^3	\$1,500,000
4.	Air piping, schedule 40, steel	0.6 m dia, 610 m	\$13/m	\$ 80,000
5.	Air piping, schedule 40, PVC	0.13 m dia, 11.700 m	\$4/m	\$ 461,000
6.	Blowers	4-450 m^3/min	\$125,000	\$ 500,000
7.	Nutrient storage facilities			\$ 14,000
8.	Nutrient make-up and feed			\$ 15,000
9.	H_2O_2 unloading and handling	150 t capacity, 100% H_2O_2		\$ 40,000
10.	H_2O_2 metering pumps	2-60 L cap.	\$8500	\$ 17,000
11.	H_2O_2 storage tanks, aluminum	250 m3		\$ 80,000
12.	Ferrous sulphate storage tank, FRP	75 m3		\$ 15,000
13.	Ferrous sulphate metering pump	2-30 L cap.	\$4500	\$ 9,000
14.	Reactor tanks, FRP, baffled	3-180 m^3	\$23,000	\$ 69,000
15.	Agitators	3-30 kW	\$40,000	\$ 120,000
16.	Building	610 m^2	\$750/m^2	\$ 455,000

OPERATING COSTS

No.	Item Description	Amount	Unit Cost	Total Cost
1.	Electric power	14.5×10^6 kW h	\$0.035 kW h	\$ 506,000
2.	Nutrients - Diammonium phosphate	70 t/y	\$1700/t	\$ 119,000
	- Ammonium phosphate	250 t/y	\$370/t	\$ 92,500
3.	Operating labour	4 my	\$50,000/my	\$ 200,000
4.	Hydrogen peroxide	1000 t/y	\$1300/t	\$1,300,000
5.	Ferrous sulphate	220 t/y	\$120/t	\$ 26,400

TABLE B-3

PVP-Cu CATALYZED AIR OXIDATION

CAPITAL COSTS

No.	Item Description	Quantity/Size	Unit Cost	Total Cost
1.	Reagent Storage/Feeding			
1.1	Copper sulphate storage bin	6 t capacity	$7500	$ 7,500
	Copper sulphate feeder	1 t/d capacity	$5000	$ 5,000
1.2	NaOH storage tank	15 m^3	$5000	$ 5,000
	NaOH metering pump		$7500	$ 7,500
	pH controls		$5000	$ 5,000
1.3	PVP resin storage bin	10 t capacity	$9000	$ 9,000
	PVP feed system	1 t/d capacity	$5000	$ 5,000
2.	Aerated Pond			
2.1	Earthen basins, concrete bottom, clay-lined walls	2		
	Cut and fill	15,500 m^3	$5.20/$m^3$	$ 80,000
	Clay liner	1150 m^3	$13/$m^3$	$ 15,000
	Concrete bottom, formless, 5 cm	56 m^3	$400/$m^3$	$ 22,500
	Submerged combustion concrete basin			$ 75,000
2.2	Insulated top cover, 5 cm styrofoam blankets	5700 m^2	$11/$m^2$	$ 61,600
2.3	Aeration System			
	Static aerators	1100	$250	$ 275,000
	Air piping, ABS or PVC	0.15 m dia, 1050 m	$48/m	$ 50,800
	Air piping, ABS or PVC	0.75 m dia, 150 m	$185/m	$ 27,500
	Blowers	3-450 m^3/min	$125,000	$ 375,000
2.4	Solid/Liquid Separation			
	Vibrascreen units, double-deck	3	$15,000	$ 45,000

TABLE B-3

PVP-Cu CATALYZED AIR OXIDATION (CONT'D)

No.	Item Description	Quantity/Size	Unit Cost	Total Cost
2.5	Catalyst Regeneration			
	Belt presses	2	$50,000	$ 100,000
	Reactor vessels, FRP	2-2.5 m^3 capacity	$4,000	$ 8,000
	Agitators	2-0.4 kW	$3,000	$ 6,000
	Reactor vessels, FRP	2-10 m^3 capacity	$5,500	$ 11,000
	Agitators	2-1.5 kW	$4,000	$ 8,000
2.6	Submerged combustion equipment	5 x 10^8 kcal/d capacity		$ 125,000
2.7	Initial catalyst charge, PVP resin	227,000 kg	$66/kg	$15,000,000
2.8	Buildings	4000 m^2	$750/$m^2$	$ 300,000

OPERATING COSTS

No.	Item Description	Amount	Unit Cost	Total Cost
1.	Copper sulphate	300 t/y	$680/t	$ 204,000
2.	Fuel for submerged combustion	50,000 M kcal/y	$15.80/M kcal	$ 800,000
3.	Electric power	13.1 x 10^6 kW h/y	$0.035/kW h	$ 460,000
4.	PVP Catalyst resin make-up	227,000 kg/y	$66/kg	$15,000,000
5.	Sodium hydroxide	550 t/y	$475/t	$ 260,000
6.	Operating labour	4 my	$50,000/my	$ 200,000

TABLE B-4

COPPER SULPHIDE CATALYZED AIR OXIDATION

CAPITAL COSTS

No.	Item Description	Quantity/Size	Unit Cost	Total Cost
1.	Reagent Storage/Preparation/Feeding			
1.1	White metal mill	30 t/d capacity	$250,000	$ 250,000
	White metal storage, feeding	100 t silo, 30 t/day feeding	$50,000	$ 50,000
1.2	Sodium hydroxide storage tank	50 m^3	$10,000	$ 10,000
	Sodium hydroxide feed pump		$7,500	$ 7,500
	pH controls		$5,000	$ 5,000
2.	Process Equipment/Units			
2.1	Earthen basins, concrete bottom, clay-lined walls	2		
	Cut and fill	31,000 m^3	$5.20/$m^3$	$ 160,000
	Clay liner	1550 m^3	$13/$m^3$	$ 20,000
	Concrete bottom, formless, 5 cm	95 m^3	$400/$m^3$	$ 37,500
	Submerged combustion concrete basin			$ 100,000
2.2	Insulated top cover, 5 cm	5900 m^2	$11/$m^2$	$ 65,000
2.3	Aeration System			
	Static aerators	1750	$250	$ 437,500
	Piping, ABS or PVC	0.15 m dia, 2175 m	$48/m	$ 103,000
	Piping, ABS or PVC	0.75 m dia, 180 m	$185/m	$ 36,500
	Blowers	4 · 735 m^3/min	$140,000	$ 560,000
2.4	Solid/Liquid Separation			
	Vibrascreen units, double-deck	3	$15,000	$ 45,000
2.5	Slurry Dewatering			
	Belt Press	1	$50,000	$ 50,000
2.6	Submerged combustion equipment	2·5 x 10^8 kcal/d capacity	$125,000	$ 250,000
2.7	Initial catalyst charge	690 t	$1390/t	$ 960,000
2.8	Building	400 m^2	$750/$m^2$	$ 300,000

OPERATING COSTS

No.	Item Description	Amount	Unit Cost	Total Cost
1.	Fuel for submerged combustion	50,000 M kcal/y	$15,80/M kcal	$ 800,000
2.	Electric power	19.7 x 10^6 kW h/y	$0.035/kW h	$ 690,000
3.	White metal make-up (loss in effluent)	100 kg/d	$1.98/kg	$ 510,000
4.	White metal reprocessing/ transportation costs	10,800 t/y	$100/t	$1,080,000
5.	Operating labour	8 my	$50,000/my	$ 400,000

TABLE B-5

HYDROGEN PEROXIDE OXIDATION

CAPITAL COSTS

No.	Item Description	Quantity/Size	Unit Cost	Total Cost
1.	Reagent Storage/Feeding			
1.1	H_2O_2 unloading and handling			$ 50,000
1.2	H_2O_2 storage tanks, aluminum	500 m3 capacity		$ 120,000
1.3	H_2O_2 metering pumps	2-60 L/min. capacity	$8,500	$ 17,000
1.4	Ferrous sulphate, storage tank, FRP	150 m3 capacity		$ 20,000
1.5	Ferrous sulphate metering pumps	2-30 L/min capacity	$4,500	$ 9,000
2.	Process Equipment			
2.1	Reactor tanks, FRP, baffled	3-100 m^3 capacity	$23,000	$ 69,000
2.2	Agitators	3-30 kW	$40,000	$ 120,000
3.	Building	560 m^2	$750/$m^2$	$ 420,000

OPERATING COSTS

No.	Item Description	Amount	Unit Cost	Total Cost
1.	H_2O_2	10,000 t/y	$1,300/t	$13,000,000
2.	Ferrous sulphate	2,200 t/y	$120/t	$ 264,000
3.	Electric power	1.97×10^6 kW hr/y	$0.035/kW h	$ 69,000
4.	Operating labour	4 my	$50,000/my	$ 200,000

TABLE B-6

CHLORINE OXIDATION

CAPITAL COSTS

No.	Item Description	Quantity/Size	Unit Cost	Total Cost
1.	Chlorine storage - directly from rail cars			
2.	Unloading and handling system (pumps, valves, piping)			$ 50,000
3.	Reactor tanks, FRP, baffled	3-180 m^3 capacity	$23,000	$ 69,000
4.	Mixers	3-30 kW	$40,000	$ 120,000
5.	Building	280 m^2	$750/$m^2$	$ 210,000

OPERATING COSTS

No.	Item Description	Amount	Unit Cost	Total Cost
1.	Chlorine	41,650 t/y	$325/T	$13,500,000
2.	Electric power	1.97×10^6 kW h	$0.035/kW h	$ 69,000
3.	Operating labour	8 my	$50,000	$ 400,000

TABLE B-7

OZONE OXIDATION

CAPITAL COSTS

No.	Item Description	Quantity/Size	Unit Cost	Total Cost
1.	Oxygen plant	30,000 kg/d capacity		\$ 3,800,000
2.	Ozone generation system and oxygen recycle (includes compressors, heat exchangers, driers, ozonators)	26,000 kg/d ozone		\$20,200,000
3.	Reactors, covered, baffled	3 · 225 m^3 capacity	\$100,000	\$ 300,000
4.	Mixers	3 · 37.5 kW	\$48,000	\$ 144,000
5.	Ozone destruction system	280 m^3 capacity		\$ 150,000
6.	Building	465 m^2	\$750/m^2	\$ 350,000

OPERATING COSTS

No.	Item Description	Amount	Unit Cost	Total Cost
1.	Electric Power		\$0.035/kW h	
1.1	Mixers	1.0×10^6 kW h/y		\$ 34,500
1.2	Ozone destruct furnace	1.0×10^7 kW h/y		\$ 345,000
1.3	Ozonators	1.0×10^8 kW h/y		\$ 3,450,000
1.4	Oxygen recycle	2.7×10^7 kW h/y		\$ 916,000
2.	Oxygen plant operation	12,000 t/y capacity	\$22/t	\$ 264,000
3.	Operating labour	8 my	\$50,000	\$ 400,000

TABLE B-8

SO_2-AIR OXIDATION

CAPITAL COSTS

No.	Item Description	Quantity/Size	Unit Cost	Total Cost
1.	Reagent Storage/Feeding Facilities			$ 50,000
1.1	Sulphur unloading, storage (250 t capacity), 110 m^3 tank c/w steam coils, tank car heating system, pumps			
1.2	Sulphur burner	50 t/d capacity		$ 25,000
1.3	Ducting and blower for SO_2 gas	30 m^3/min, 35 kW		$ 94,300
1.4	$CuSO_4.5H_2O$ storage bin (10t), feeder (1.5 t/d)			$ 22,100
2.	Process Equipment			
2.1	Air blowers	3 - 420 m^3/min	$75,000	$ 225,000
2.2	Reactor tanks, FRP, baffled	6 - 180 m^3 capacity	$23,000	$ 138,000
2.3	Mixers	6 - 30 kW	$40,000	$ 240,000
2.4	Air ducting, FRP	1.2 dia, 45 m	$650/m	$ 30,000

OPERATING COSTS

No.	Item Description	Amount	Unit Cost	Total Cost
1.	Sulphur	13,000 t/y	$150/t	$ 1,950,000
2.	Copper sulphate	500 t/y	$680/t	$ 340,000
3.	Electric power	16.5 x 10^6 kW hr/y	$0.035/kW h	$ 575,000
4.	Operating labour	2 my	$50,000	$ 100,000

TABLE B-9

ELECTROCHEMICAL OXIDATION

CAPITAL COSTS

No.	Item Description	Quantity/Size	Unit Cost	Total Cost
1.	Sand filter	75 m^2 area		$ 1,000,000
2.	Electrochemical facility	9000 m^2 area, 20 groups of 6 cells	$2,000,000	$40,000,000

OPERATING COSTS

No.	Item Description	Amount	Unit Cost	Total Cost
1.	Electrical power	294 x 10^6 kW h	$0.035/kW h	$10,300,000
2.	Electrical replacement costs			$ 2,000,000
3.	Operating labour	4 my	$50,000/my	$ 200,000

TABLE B-10

REVERSE OSMOSIS AND EVAPORATION

CAPITAL COSTS

No.	Item Description	Quantity/Size	Unit Cost	Total Cost
1.	Sand filter	75 m^2		\$ 1,000,000*
2.	Phosphonate, storage & feeding			\$ 30,800
3.	Reverse osmosis, modular system, including high pressure pumps	720 m^3/h capacity		\$ 6,200,000
4.	Evaporators, long-tube, vertical	1,400 m^2 surface area		\$ 900,000
5.	Building	370 m^2	\$750/m^2	\$ 280,000

* Installed

OPERATING COSTS

No.	Item Description	Amount	Unit Cost	Total Cost
1.	Power	56 million kW h	\$.035/kW h	\$ 1,962,000
2.	Chemicals		\$.033/m^3 water	\$ 263,000
3.	Operation	8 my	\$50,000/y	\$ 400,000
4.	Membrane replacement, US gal.		\$0.33/m^3 water	\$ 2,628,000
5.	Evaporation		\$2,20/m^3 water	\$ 4,380,000

TABLE B-11

REVERSE OSMOSIS AND ALKALINE PRESSURE OXIDATION

CAPITAL COSTS

No.	Item Description	Quantity/Size	Unit Cost	Total Cost
1.	Sand filter	75 m^2		$ 1,000,000*
2.	Phosphonate, storage & feeding			$ 30,000
3.	Reverse Osmosis, modular system, (including high pressure pumps)	720 m^3/h		$ 6,200,000*
4.	Alkaline pressure oxidation module (including heat exchanger autoclave, steam plant)			$ 3,300,000*
5.	Oxygen plant	30 t/d capacity		$ 3,600,000
6.	Building	370 m^2	$\$750/\mathrm{m}^2$	

* Installed

OPERATING COSTS

No.	Item Description	Amount	Unit Cost	Total Cost
1.	Power			
1.1	Reverse osmosis	56 million kW h	$.035/kW h	$1,962,000
1.2	Power for agitators, pumps	0.98 million kW h		$ 34,500
2.	Chemicals		$\$0.033/\mathrm{m}^3$ water	$ 263,000
3.	Operation	8 my	$50,000/y	$ 400,000
4.	Membrane replacement		$\$0.33/\mathrm{m}^3$ water	$2,628,000
5.	Caustic	3650 t/y	$400/t	$1,460,000
6.	Steam	0.15 million t/y	$15/t	$2,240,000
7.	Oxygen	8500 t/y	$22/t	$ 187,000

TABLE B-12

REVERSE OSMOSIS AND ELECTROCHEMICAL

CAPITAL COSTS

No.	Item Description	Quantity/Size	Unit Cost	Total Cost
1.	Sand filter	75 m^2		$ 1,000,000*
2.	Phosphonate, storage & feeding			$ 30,000
3.	Reverse osmosis, modular system, (including high pressure pumps)			$ 6,200,000*
4.	Electrochemical thiosalt oxidation			
4.1	Electrochemical reactors	4500 m^2, 10 groups of six cells		$13,000,000*
5.	Building	280 m^2	$\$750/\mathrm{m}^2$	$ 210,000

* Installed

OPERATING COSTS

No.	Item Description	Amount	Unit Cost	Total Cost
1.	Power			
1.1	Reverse osmosis	56 million kW h	$.035/kW h	$1,962,000
1.2	Electrochemical		$.035/kW h	$3,400,000
2.	Chemicals		$\$0.033/\mathrm{m}^3$ water	$ 263,000
3.	Operation	8 my	$50,000/y	$ 400,000
4.	Membrane replacement		$\$0.33/\mathrm{m}^3$ water	$2,628,000
5.	Electrode replacement			$ 700,000

TABLE B-13

ALKALINE PRESSURE OXIDATION

CAPITAL COSTS

No.	Item Description	Quantity/Size	Unit Cost	Total Cost
1.	Heat exchangers, shell CS tubes 304SS	2 - 1000 m^2 surface each	$500,000	$ 1,000,000
2.	Auto claves	2 - 180 m^3	$960,000	$ 1,920,000
3.	Agitators	2 - 30 kW	$40,000	$ 80,000
4.	NaOH storage tank	100 m^3		$ 15,000
5.	NaOH metering pumps	10 t/d capacity		$ 10,000
6.	Steam plant	1.8 million kg/d capacity		$ 1,000,000
7.	Oxygen plant	30 t/d capacity		$ 3,600,000*
8.	Building (excluding steam & oxygen plants)	210 m^2	$720/m^2	$ 160,000

OPERATING COSTS

No.	Item Description	Amount	Unit Cost	Total Cost
1.	Steam	5.9 x 10^8 kg/y	$15./1000 kg	$ 8,900,000
2.	Electric power	3.4 x 10^6 kW h/y	$0.035/kW h	$ 120,000
3.	Sodium hydroxide	3650 t/y	$400/t	$ 1,460,000
4.	Oxygen generation	8500 t/y	$22/t	$ 187,000

* installed cost

TABLE B-14

REDUCTION BY IRON

CAPITAL COSTS

No.	Item Description	Quantity/Size	Unit Cost	Total Cost
1.	Heat exchangers, surface area	2 - 1000 m^2	$500,000	$ 1,000,000
2.	Covered tanks	4 - 350 m^3	$150,000	$ 600,000
3.	Agitators	4 - 110 kW	$100,000	$ 400,000
4.	Iron powder, make-up, storage and handling system	54 t/d capacity		$ 400,000
5.	Hydrogen sulphide venting (blower & piping)			$ 45,000
6.	Steam plant	1.8 million kg/d capacity		$ 1,000,000
7.	Hydrogen peroxide storage	20 t/capacity		$ 20,000
8.	Hydrogen peroxide feeding	2 t/d capacity		$ 10,000
9.	Hydrogen sulphide reactor & agitator	5 m^3 capacity		$ 15,000
10.	Sulphur filter			$ 50,000
11.	Building	280 m^2	$750/$m^2$	$ 210,000

OPERATING COSTS

No.	Item Description	Amount	Unit Cost	Total Cost
1.	Steam	5.8 x 10^8 kg/y	$15/1000 kg	$ 8.940,000
2.	Iron pellets	20,000 t/y	$200/t	$ 4,000,000
3.	Power	6.5 x 10^6 kW h/y	$0.035/kW h	$ 230,000
4.	Hydrogen peroxide	730 t/y	$1300/t	$ 949,000
5.	Operating labour	4 my	$50,000/my	$ 200,000

TABLE B-15

ACTIVATED CARBON ADSORPTION AND THERMAL REGENERATION

CAPITAL COSTS

No.	Item Description	Quantity/Size	Unit Cost	Total Cost
1.	Contactor system (4 columns in series, total contact time - 40 min.), piping, valves, carbon storage tanks, instrumentation			\$ 1,500,000*
2.	Carbon inventory	540 t	\$1.45/kg	\$ 780,000
3.	Thermal Regeneration			
3.1	Feeding and conveying system, multiple hearth furnace	1900 m^2 hearth area		\$ 2,000,000*
3.2	Afterburner, scrubber, water treatment system (including reactor, clarifier, lime storage facilities, sludge disposal system)			\$ 6,000,000
3.3	Carbon storage, dewatering, building			\$ 450,000*
3.4	Steam boiler	270 t/d capacity		\$ 265,000

* Installed

OPERATING COSTS

No.	Item Description	Amount	Unit Cost	Total Cost
1.	Activated carbon make up	8,800 t/y	\$1.45/kg	\$12,800,000
2.	Carbon regeneration	270 t/d	\$0.55/kg	\$49,200,000
3.	Scrubber water neutralization			\$ 1,000,000
4.	Power costs	3.5×10^6 kW h	\$0.035/kW h	\$ 120,000
5.	Operating labour	8 my	\$50,000	\$ 400,000

TABLE B-16

ACTIVATED CARBON ADSORPTION - ELECTROCHEMICAL OXIDATION

CAPITAL COSTS

No.	Item Description	Quantity/Size	Unit Cost	Total Cost
1.	Contactor system (4 columns in series, total conact time - 40 min.), piping, valves, carbon storage tanks, instrumentation			\$ 1,500,000*
2.	Carbon inventory	540 t activated carbon	\$1.45/kg	\$ 780,000
3.	Chemical Regeneration			
3.1	Caustic storage and feeding	450 m^3		\$ 100,000
3.2	Regeneration reactor & agitator	110 m^3, 40 kW		\$ 60,000
3.3	Vibra-screen			\$ 15,000
3.4	Carbon storage, dewatering building			\$ 450,000*
4.	Electrochemical Thiosalt Oxidation			
4.1	Electrochemical reactors	1800 m^2 area, 4 groups of 6 cells		\$ 8,000,000*
5.	Building	185 m^2	\$750/m^2	\$ 140,000

* Installed

OPERATING COSTS

No.	Item Description	Amount	Unit Cost	Total Cost
1.	Activated carbon make-up	890 t/y	\$1.45/kg	\$ 1,300,000
2.	Caustic	25 t/d	\$400/t	\$ 3,600,000
3.	Power	60 x 10^6 kW h/y	\$0.035/kW h	\$ 2,100,000
4.	Lime, unslaked	3000 t/y	\$100/t	\$ 300,000
5.	Operating labour	8 my/y	\$50,000/my	\$ 400,000

TABLE B-17

ACTIVATED CARBON ADSORPTION - ALKALINE PRESSURE OXIDATION

CAPITAL COSTS

No.	Item Description	Quantity/Size	Unit Cost	Total Cost
1.	Contactor system (4 columns in series, total contact time - 40 min.), piping, valves, carbon storage tanks, instrumentation			$ 1,500,000*
2.	Carbon inventory	540 t	$1.45 kg	$ 780,000
3.	Chemical Regeneration			
3.1	Caustic storage feeding	450 m^3		$ 100,000
3.2	Regeneration reactor, & agitator	110 m^3, 40 kW		$ 60,000
3.3	Vibra-screen			$ 15,000
3.4	Carbon storage, dewatering building			$ 450,000*
4.	Alkaline pressure oxidation (including heat exchanger, autoclave, steam plant)			$ 1,800,000*
5.	Oxygen plant	30 t/d capacity		$ 3,600,000
6.	Building	190 m^2	$750/$m^2$	$ 140,000

* Installed

OPERATING COSTS

No.	Item Description	Amount	Unit Cost	Total Cost
1.	Activated carbon make up	890 t/y	$1.45/kg	$ 1,300,000
2.	Caustic	25 t/d	$400/t	$ 3,600,000
3.	Steam	6×10^7 kg/y	$15/1000 kg	$ 900,000
4.	Lime, unslaked	3,000 t	$100/t	$ 300,000
5.	Labour	8 my/y	$50,000/y	$ 400,000
7.	Oxygen	8500 t/y	$22/t	$ 187,000

TABLE B-18

DISPOSAL TO THE SEA

CAPITAL COSTS

No.	Item Description	Quantity/Size	Unit Cost	Total Cost
1.	Pipe, PVC, including connection, fusing, installation and burial costs	0.5 m dia, 34 km	\$300/m	\$10,200,000*
2.	Road crossings	6	\$10,000	\$ 60,000
3.	Right of way costs			\$ 2,500,000
4.	Pumping System			
4.1	Pumps	2 - 300 HP	\$150,000	\$ 300,000
4.2	Air compressor			\$ 25,000
4.3	After cooler			\$ 15,000
4.4	Intake system			\$ 40,000
5.	Building	53 m^2	\$750/m^2	\$ 435,000
6.	Diffuser	1.5 km		\$ 900,000*

* Installed cost

OPERATING COSTS

No.	Item Description	Amount	Unit Cost	Total Cost
1.	Electric power	2×10^6 kW h/y	\$0.035/kW h	\$ 70,000